U.S. DEPARTMENT OF COMMERCE

PB80-173636

# Technology Assessment Report for Industrial Boiler Applications: NOx Flue Gas Treatment

Radian Corp.
Austin, TX

Prepared for

Industrial Environmental Research Lab.
Research Triangle Park, NC

Dec 79

**U.S. Department of Commerce**
National Technical Information Service
NTIS

United States
Environmental Protection
Agency

Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

EPA-600/7-79-178g
December 1979

PB80-173636

EPA

# Technology Assessment Report for Industrial Boiler Applications: $NO_x$ Flue Gas Treatment

## Interagency Energy/Environment R&D Program Report

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## TECHNICAL REPORT DATA

*(Please read Instructions on the reverse before completing)*

| | | |
|---|---|---|
| 1. REPORT NO.<br>EPA-600/7-79-178g | 2. | 3. RECIPIENT'S ACCESSION NO.<br>PB80-173636 |
| 4. TITLE AND SUBTITLE<br>Technology Assessment Report for Industrial Boiler Applications: NOx Flue Gas Treatment | | 5. REPORT DATE<br>December 1979<br>6. PERFORMING ORGANIZATION CODE |
| 7. AUTHOR(S)<br>Gary D. Jones and Kevin L. Johnson | | 8. PERFORMING ORGANIZATION REPORT NO. |
| 9. PERFORMING ORGANIZATION NAME AND ADDRESS<br>Radian Corporation<br>8500 Shoal Creek Boulevard<br>Austin, Texas 78766 | | 10. PROGRAM ELEMENT NO.<br>INE624<br>11. CONTRACT/GRANT NO.<br>68-02-2608, Task 45 |
| 12. SPONSORING AGENCY NAME AND ADDRESS<br>EPA, Office of Research and Development<br>Industrial Environmental Research Laboratory<br>Research Triangle Park, NC 27711 | | 13. TYPE OF REPORT AND PERIOD COVERED<br>Task Final; 6/78 - 11/79<br>14. SPONSORING AGENCY CODE<br>EPA/600/13 |

15. SUPPLEMENTARY NOTES IERL-RTP project officer is J. David Mobley, Mail Drop 61, 919/541-2915.

16. ABSTRACT The report gives results of an assessment of the applicability of NOx flue gas treatment (FGT) technology to industrial boilers and is one of a series of technology assessment reports to aid in determining the technological basis for a New Source Performance Standard for Industrial Boilers. The status of development and performance of alternative NOx FGT control techniques were assessed and the cost, energy, and environmental impacts of the most promising processes were identified. It was found that processes utilizing selective catalytic reduction (SCR) of NOx with ammonia can achieve 90% reduction of NOx emissions, and that these processes are the nearest to commercialization in the U.S. Cost estimates of applying SCR processes in the U.S. indicated that the cost effectiveness varies significantly depending on the fuel fired, boiler size, and control level. However, boiler size is the most significant factor affecting cost effectiveness with the economy of scale causing control of large sources to be the most effective. The energy impact of applying SCR processes averaged about 0.5% of boiler capacity. No adverse environmental impacts were apparent although emissions, liquid effluents, and solid wastes must be controlled. For regulatory purposes this assessment must be viewed as preliminary, pending results of the more extensive impact studies required by Clean Air Act Sect. 111.

17. KEY WORDS AND DOCUMENT ANALYSIS

| a. DESCRIPTORS | | b. IDENTIFIERS/OPEN ENDED TERMS | c. COSATI Field/Group | |
|---|---|---|---|---|
| Pollution | Catalysts | Pollution Control | 13B | 07D |
| Nitrogen Oxides | Ammonia | Stationary Sources | 07B | |
| Flue Gases | Fossil Fuels | Flue Gas Treatment | 21B | 21D |
| Assessments | | Selective Catalytic Reduction (SCR) | 14B | |
| Industrial Processes | | | 13H | |
| Boilers | | | 13A | |

| | | |
|---|---|---|
| 18. DISTRIBUTION STATEMENT<br>Release to Public | 19. SECURITY CLASS (This Report)<br>Unclassified | 21. NO. OF PAGES |
| | 20. SECURITY CLASS (This page)<br>Unclassified | 22. PRICE |

EPA Form 2220-1 (9-73)

EPA-600/7-79-178g

December 1979

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by

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Contract No. 68-02-2608
Task No. 45
Program Element No. INE624

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Office of Research and Development
Washington, DC 20460

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# ABSTRACT

This study assesses the applicability of $NO_x$ flue gas treatment technology to industrial boilers and is one of a series of technology assessment reports to aid in determining the technological basis for a New Source Performance Standard for Industrial Boilers. The status of development and performance of alternative $NO_x$ flue gas treatment control techniques were assessed and the cost, energy, and environmental impacts of the most promising processes were identified. It was found that processes utilizing selective catalytic reduction (SCR) of $NO_x$ with ammonia can achieve 90 percent reduction of $NO_x$ emissions, and that these processes are the nearest to commercialization in the U.S. In Japan, SCR processes have been successfully operated on commercial scale gas-and oil-fired sources and are being installed on coal-fired sources. Cost estimates of applying SCR processes in the U.S. indicated that the cost effectiveness varies significantly depending on the fuel fired, boiler size, and control level. However, boiler size is the most significant factor affecting cost effectiveness with the economy of scale causing control of large sources to be the most effective. The energy impact of applying SCR processes averaged about 0.5 percent of boiler capacity. No adverse environmental impacts were apparent although there are emissions, liquid effluents, and solid wastes that must be controlled. For regulatory purposes this assessment must be viewed as preliminary, pending the results of the more extensive examination of impacts called for under Section 111 of the Clean Air Act.

# PREFACE

The 1977 Amendments to the Clean Air Act required that emission standards be developed for fossil-fuel-fired steam generators. Accordingly, the U.S. Environmental Protection Agency (EPA) recently promulgated revisions to the 1971 new source performance standard (NSPS) for electric utility steam generating units. Further, EPA has undertaken a study of industrial boilers with the intent of proposing an NSPS for this category of sources. The study is being directed by EPA's Office of Air Quality Planning and Standards, and technical support is being provided by EPA's Office of Research and Development. As part of this support, the Industrial Environmental Research Laboratory at Research Triangle Park, NC, prepared a series of technology assessment reports to aid in determining the technological basis for the NSPS for industrial boilers. This report is part of that series. The complete report series is listed below:

| Title | Report No. |
|---|---|
| The Population and Characteristics of Industrial/ Commercial Boilers | EPA-600/7/79-178a |
| Technology Assessment Report for Industrial Boiler Applications: Oil Cleaning | EPA-600/7-79-178b |
| Technology Assessment Report for Industrial Boiler Applications: Coal Cleaning and Low Sulfur Coal | EPA-600/7-79-178c |
| Technology Assessment Report for Industrial Boiler Applications: Synthetic Fuels | EPA-600/7-79-178d |
| Technology Assessment Report for Industrial Boiler Applications: Fluidized-Bed Combustion | EPA-600/7-79-178e |
| Technology Assessment Report for Industrial Boiler Applications: $NO_x$ Combustion Modification | EPA-600/7/79-178f |

| Title | Report No. |
|---|---|
| Technology Assessment Report for Industrial Boiler Applications: $NO_x$ Flue Gas Treatment | EPA-600/7-79-178g |
| Technology Assessment Report for Industrial Boiler Applications: Particulate Collection | EPA-600/7-79-178h |
| Technology Assessment Report for Industrial Boiler Applications: Flue Gas Desulfurization | EPA-600/7-79-178i |

These reports will be integrated along with other information in the document, "Industrial Boilers - Background Information for Proposed Standards," which will be issued by the Office of Air Quality Planning and Standards.

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# ACKNOWLEDGEMENTS

This report would not have been possible without the assistance of several people. The authors would like to express their appreciation to the following people for their support in the preparation of this report.

- The process vendors who supplied much of the data used for the analyses.
- Dr. Jumpei Ando for supplying information concerning $NO_x$ flue gas treatment systems in Japan.
- C. B. Sedman and L. D. Broz for their coordination efforts throughout the program.
- J. D. Mobley for his guidance and assistance.
- M. Harris, J. C. Fischer, and C. K. Holcomb for their work in typing this report.

# SECTION 1

# EXECUTIVE SUMMARY

## 1.1 INTRODUCTION

### 1.1.1 Background and Objectives

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency to coordinate and lead the development and implementation of regulations on air pollution. These include standards of performance for new and modified sources of pollution. Fossil fired steam generators are specifically mentioned in the act and EPA has undertaken a study of industrial boilers with intent to propose emission control levels based upon the results of this and other studies.

This specific report examines the impacts of application of flue gas treatment (FGT) for $NO_x$ control on industrial boilers. The overall objective is to provide a background document that quantifies the economic, energy and environmental impacts as well as establish whether or not the technology is demonstrated and available to the U.S. market. All potential FGT technologies are considered and detailed analyses are performed on those which are the most promising.

### 1.1.2 Report Organization and Approach

Several boiler/FGT combinations are considered in the detailed analyses that follow. In Section 2, all $NO_x$ control processes that have been developed to treat boiler flue gas are discussed in moderate detail. The section is divided into three subsections based on fuel--coal, oil, and gas. This is done for two reasons: 1) to make this report consistent in format with other Individual Technology Assessment Reports (ITAR's, see preface), and 2) to examine the effect of fuel type on the various technologies considered. In the case of FGT, the majority of the technologies can be applied to the majority of the fuels. As a result, much of the material in the three subsections is very similar, especially with regard to the technical descriptions of the systems.

A decision was made early in the development of this report to produce essentially three stand-alone sections; one for each fuel type. This allows one or more fuel types to be eliminated from consideration without impacting the quality of the data in the remaining sections. As a result, there is a significant amount of repetition in the three subsections. The subsection dealing with applications to oil-fired boilers (Section 2.3) contains descriptions of all of the FGT technologies considered and has the greatest amount of information on specific systems. Therefore, for most readers, it is necessary to read only subsections 2.1 and 2.3 for a complete description of all FGT technologies considered. Subsections 2.2 and 2.4, dealing with coal- and gas-fired applications, do contain unique information on the status of development and number of applications and can be consulted if this specific information is desired. The Executive Summary is organized differently than the body of the report in that all fuel types are discussed together instead of separately. This is done to allow the reader to directly examine the effect of fuel type on the economic, energy, and environmental impacts. The summary discusses each of the impacts separately and also separates $NO_x$-only systems from $NO_x/SO_x$ systems.

The large number of potential fuel/boiler/FGT system combinations requires the combinations be reduced to those systems with a high potential for commercial application and successful operation. This is done in Section 3. The data used to make these preliminary evaluations is derived from Section 2. The combinations selected in Section 3 are then analyzed in detail in subsequent sections.

Section 4 presents the economic impacts of these FGT processes on ten industrial boilers. Standard costing techniques are used to develop annualized costs which are plotted to show the effect of several parameters on the total system costs. The process specifications used in the economic analyses are developed in Sections 5 and 6. These two Sections present the results of material and energy balances performed for each case to quantify the energy and environmental impacts of FGT systems. These balances are also used to size most of the individual pieces of equipment.

In Sections 4, 5 and 6 results are not presented for all possible control levels. That is, while some systems have data presented for three levels of control, others have data presented for only two or one control level(s). The curves, however, are plotted over the range of the three control levels (70, 80 and 90 percent). When the individual Sections were initially prepared, data was calculated for all three control levels. During the interim period prior to the compilation of these Sections into the final report, new cases were added and several economic premises were changed. In order to meet budget and time constraints, it was necessary to reduce the number of analyses.

It was observed that curve shapes were all very similar and that new curves could be drawn accurately without a complete set of points. In cases with a lot of similarity, i.e. among the coal-fired boilers, a curve shape was established for one case by using three points. For the other cases, an analysis was made to determine the midpoint of the curve. Curves, similar in shape to that developed by a three-point analysis, were then passed through these midpoints. In cases where there was little similarity, a two-point analysis was performed to determine the end-points of the curve. A curve was then drawn through these points using the original curve (determined by three-point analysis in the initial case analysis) to determine the shape. It should be noted that, even if straight lines were used, the interpolated and extrapolated results would not be changed significantly. This is why a limited set of analyses are used to determine a complete set of cost and energy data.

The final section, Section 7, deals with test data that have been determined for operating FGT systems. These types of data do not exist for U.S. applications since FGT has yet to be applied in this country. A Japanese consultant with contacts among FGT system users was retained to obtain test data from industrial boilers in Japan. The test data presented in Section 7 represents the most complete set of data of this type available.

For the reader interested in the details involved with the analyses presented in this report, the Appendices present an example calculation as well as complete sets of process selection criteria material and energy balances, and cost breakdowns.

### 1.1.3 Scope of Study

Several variables are considered in order to make the study as comprehensive as possible, these being

- Fuel
- Boiler, type and size
- Control level
- FGT process type
- Pollutants controlled.

As mentioned previously, three fuel types are considered: coal, oil and natural gas. Coal and oil are further divided as shown below.

Coal

High Sulfur Eastern (3.5% S)
Low Sulfur Eastern (0.9% S)
Low Sulfur Western (0.6% S)

Oil

Distillate
Residual

One boiler type is considered for natural gas, distillate oil and residual oil. However, four boiler types are considered for the coal fuels. The combinations of fuel and boiler type considered at the beginning of the study are shown in Table 1.1.3-1. These boilers are termed "standard boilers" because they apply to all of the ITAR's. The $NO_x$ emissions from these boilers are shown in Table 1.1.3-2.

In the ensuing discussion of emission control technologies, candidate technologies were compared using three emission control levels labeled "moderate, intermediate, and stringent." These control levels were chosen only to encompass all candidate technologies and form bases for comparison of technologies for control of specific pollutants considering performance, costs, energy, and non-air environmental effects.

From these comparisons, candidate "best" technologies for control of individual pollutants are recommended by the contractor for consideration in subsequent industrial boiler studies. These "best technology" recommendations do not consider combinations of technologies to remove all pollutants and have not undergone the detailed environmental, cost, and energy impact assessments necessary for regulatory action. Therefore, the levels of "moderate, intermediate, and stringent" and the recommendation of "best technology" for individual pollutants are not to be construed as indicative of the regulations that will be developed for industrial boilers. EPA will

TABLE 1.1.3-1. CHARACTERISTICS OF THE STANDARD BOILERS

| Type | Fuel* | Rating $MW_t$ (MBtu/hr) | Gas Flow Rate $Nm^3$/hr |
|---|---|---|---|
| Package, Firetube | Distillate Oil | 4.4 (15) | 5,400 |
| Package, Firetube | Natural Gas | 4.4 (15) | 5,600 |
| Package, Watertube | Residual Oil | 8.8 (30) | 9,500 |
| Package, Watertube<br>Underfeed Stoker | HSE | 8.8 (30) | 12,500 |
| | LSE | 8.8 (30) | 12,600 |
| | LSW | 8.8 (30) | 12,900 |
| Package, Watertube<br>Chaingrate Stoker | HSE | 22 (75) | 31,300 |
| | LSE | 22 (75) | 31,000 |
| | LSW | 22 (75) | 32,400 |
| Package, Watertube | Natural Gas | 44 (150) | 52,800 |
| Package Watertube | Distillate Oil | 44 (150) | 51,900 |
| Package, Watertube | Residual Oil | 44 (150) | 47,800 |
| Field Erected, Watertube<br>Spreader Stoker | HSE | 44 (150) | 62,900 |
| | LSE | 44 (150) | 62,700 |
| | LSW | 44 (150) | 64,700 |
| Field Erected, Watertube<br>Pulverized Coal | HSE | 58.6 (200) | 72,600 |
| | LSE | 58.6 (200) | 72,800 |
| | LSW | 58.6 (200) | 75,500 |

*Coal types: HSE = High Sulfur Eastern (3.5% S)
LSE = Low Sulfur Eastern (0.9% S)
LSW = Low Sulfur Western (0.6% S)

### TABLE 1.1.3-2. $NO_x$ EMISSION RATES FOR THE STANDARD BOILERS

| Boiler | Fuel* | $NO_x$ Emissions | | | | |
|---|---|---|---|---|---|---|
| | | g/s | (lb/hr) | ng/J | (lb/$10^6$ Btu) | ppm |
| Package, Firetube | Distillate Oil | 0.300 | (2.38) | 68.8 | (0.16) | 97 |
| Package, Firetube | Natural Gas | 0.332 | (2.63) | 77.4 | (0.18) | 104 |
| Package, Watertube | Residual Oil | 2.02 | (16.0) | 228 | (0.53) | 373 |
| Package, Watertube Underfeed Stoker | HSE | 2.40 | (10.05) | 275 | (0.64) | 335 |
| | LSE | 2.06 | (16.35) | 237 | (0.55) | 288 |
| | LSW | 2.95 | (23.40) | 335 | (0.78) | 402 |
| Package, Watertube Chaingrate | HSE | 6.02 | (47.70) | 275 | (0.64) | 336 |
| | LSE | 5.15 | (40.80) | 232 | (0.54) | 290 |
| | LSW | 7.40 | (58.65) | 335 | (0.78) | 401 |
| Package, Watertube | Natural Gas | 3.31 | (26.26) | 75.3 | (0.18) | 110 |
| Package, Watertube | Distillate Oil | 2.99 | (23.76) | 68.0 | (0.16) | 101 |
| Package, Watertube | Residual Oil | 7.47 | (60.00) | 172 | (0.40) | 292 |
| Field Erected, Watertube Spreader Stoker | HSE | 12.0 | (95.40) | 275 | (0.64) | 337 |
| | LSE | 10.3 | (81.45) | 232 | (0.54) | 288 |
| | LSW | 14.8 | (117.15) | 335 | (0.78) | 400 |
| Field Erected, Watertube Pulverized Coal | HSE | 19.2 | (152.46) | 327 | (0.76) | 466 |
| | LSE | 16.5 | (130.50) | 280 | (0.65) | 396 |
| | LSW | 23.7 | (187.56) | 404 | (0.94) | 550 |

*Coal types: HSE = High Sulfur Eastern (3.5% S)
LSE = Low Sulfur Eastern (0.9% S)
LSW = Low Sulfur Western (0.6% S)

perform rigorous examination of several comprehensive regulatory options before any decisions are made regarding the standards for emission from industrial boilers. The control levels are defined in Table 1.1.1-3.

The types of FGT systems considered are different for each fuel type and these are discussed in subsequent sections for each specific fuel. The project schedule required that the number of potential combinations of boiler, fuel, and control level be reduced in order to keep the number of required analyses manageable. Detailed analyses were performed on the cases shown in Table 1.1.3-4. Note that these are the cases for FGT processes which remove only $NO_x$. For FGT processes which remove both $NO_x$ and $SO_x$ a separate set of cases was developed and is shown in Table 1.1.3-5.

Only one coal is considered for the $NO_x$-only cases. This is due to the fact that FGT process designs and impacts are not significantly affected by fuel sulfur content and therefore analyzing each coal type would not yield any additional information. The flue gas flow rates and $NO_x$ concentrations vary somewhat among the coal types considered, but not enough to cause much difference in the size of the necessary FGT process. With all of the FGT systems analyzed, the equipment size is primarily a function of the flue gas flow rate and secondarily a function of $NO_x$ concentration. However, since coal sulfur level can affect the environmental impact, two coal types are considered in this section.

For the processes which remove both $NO_x$ and $SO_x$, two coals are analyzed to show the effect of coal sulfur level on the various impacts. High sulfur eastern and low sulfur western were selected in order to have the widest range of coal sulfur levels. Also, $NO_x/SO_x$ processes for oil-fired boiler application are considered only for the case of residual oil since this oil has the most significant $SO_x$ emissions. $NO_x/SO_x$ processes are examined for application to these boilers to enable comparison between a simultaneous $NO_x/SO_x$ system and a combination of a $NO_x$-only system and an FGD system. This comparison will be made during a future phase of the industrial boiler evaluation, but not in this report.

## TABLE 1.1.1-3. $NO_x$ CONTROL LEVELS

| Fuel | Baseline $NO_x$ Emissions ng/J | Baseline $NO_x$ Emissions (lb/MBtu) | Level of Control: Moderate, 70% ng/J | Moderate, 70% (lb/MBtu) | Intermediate, 80% ng/J | Intermediate, 80% (lb/MBtu) | Stringent, 90% ng/J | Stringent, 90% (lb/MBtu) |
|---|---|---|---|---|---|---|---|---|
| Pulverized Coal | 404 | (0.94) | 121 | (0.28) | 80.8 | (0.19) | 40.4 | (0.094) |
| Stoker Coal | 335 | (0.78) | 101 | (0.23) | 67.0 | (0.16) | 33.5 | (0.078) |
| Residual Oil | 172 | (0.40) | 51.6 | (0.12) | 34.4 | (0.080) | 17.2 | (0.040 |
| Distillate Oil | 68 | (0.16) | 20.4 | (0.047) | 13.6 | (0.032) | 6.8 | (0.016) |
| Natural Gas | 75 | (0.18) | 22.6 | (0.053) | 15.1 | (0.035) | 7.5 | (0.018) |

Where emissions are dependent on boiler size, the largest boiler is shown.

TABLE 1.1.3-4. CASES SELECTED FOR DETAILED ANALYSIS - $NO_x$-ONLY FGT PROCESSES

| Boiler | Fuel* | Size $MW_t$ | Control Level % |
|---|---|---|---|
| Package, Firetube | Distillate Oil | 4.4 | 70, 90 |
| Package, Firetube | Natural Gas | 4.4 | 70, 90 |
| Package, Watertube | Residual Oil | 8.8 | 70, 90 |
| Package, Watertube Underfeed Stoker | LSW | 8.8 | 80 |
| Package, Watertube Chaingrate Stoker | LSW | 22 | 70, 80, 90 |
| Package, Watertube | Natural Gas | 44 | 70, 90 |
| Package, Watertube | Distillate Oil | 44 | 70, 90 |
| Package, Watertube | Residual Oil | 44 | 70, 90 |
| Field Erected, Watertube Spreader Stoker | LSW | 44 | 80 |
| Field Erected, Watertube Pulverized Coal | LSW | 58.6 | 70, 90 |

*LSW = Low Sulfur Western Coal (0.6%S)

TABLE 1.1.3-5. CASES SELECTED FOR DETAILED ANALYSIS - $SO_x/NO_x$ FGT PROCESSES

| Boiler | Fuel* | Boiler Size, $MW_t$ | Control Level % $NO_x$ | Control Level % $SO_x$ |
|---|---|---|---|---|
| Package, Watertube | Residual Oil | 44 | 80 | 85 |
| Package, Watertube Underfeed Stoker | HSE<br>LSW | 8.8 | 80 | 85 |
| Field Erected, Watertube Pulverized Coal | HSE<br>LSW | 58.6 | 80 | 85 |

*HSE = High Sulfur Eastern Coal (3.5% S)
LSW = Low Sulfur Western Coal (0.6% S)

It should be noted that FGT technology for $NO_x$ control has not yet been commercially applied to coal-fired boilers. However, pilot units have been tested and two full scale systems are scheduled. Coal-fired applications are considered here since they are currently being offered in the U.S. and it is felt that they will be demonstrated commercially in the near future.

## 1.2 FLUE GAS TREATMENT FOR CONTROL OF $NO_x$ ONLY

The systems of emission reduction considered in this study for applications to coal-fired boilers are divided into two general categories: those which remove only $NO_x$ and those which remove both $NO_x$ and $SO_x$. Here and throughout the study these two types of systems are considered separately to avoid confusion.

### 1.2.1 System Descriptions

The $NO_x$-only systems considered are as follows:

- Fixed Packed Bed Selective Catalytic Reduction (SCR)
- Moving Bed SCR
- Parallel Flow SCR
- Absorption-Oxidation

From the comparison evaluation of these systems, the candidates for "best" emission control systems were selected. These candidate systems are shown, along with a brief description, in Table 1.2.1-1.

SCR systems utilize ammonia to selectively reduce nitrogen oxides. The chemical mechanisms can be summarized by the following gas-phase reactions.

$$4NO + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \qquad (1\text{-}1)$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \qquad (1\text{-}2)$$

TABLE 1.2.1-1. CANDIDATES FOR BEST EMISSION CONTROL SYSTEM

| Process | Description | Fuel Application |
|---|---|---|
| Moving Bed SCR | Utilizes $NH_3$ to selectively reduce $NO_x$ to $N_2$; capable of achieving stringent $NO_x$ control level; catalyst (rings or pellets) gravity-bed, mechanically-screened, and returned to reactor. | Residual Oil |
| Parallel Flow SCR | Utilizes $NH_3$ to selectively reduce $NO_x$ to $N_2$; capable of achieving stringent $NO_x$ control level; special catalyst arrangement (honeycomb, parallel plate or tubes) greatly reduces particulate impaction as gas flow is parallel to catalyst surface. | Coal<br>Residual Oil |
| Fixed Packed Bed SCR | Utilizes $NH_3$ to selectively reduce $NO_x$ to $N_2$; capable of achieving stringent $NO_x$ control level; ring shaped catalyst pellets packed in fixed bed. | Distillate Oil<br>Natural Gas |

The first reaction predominates as flue gas $NO_x$ consists primarily of NO. Oxygen is in large excess in the flue gas and does not limit the extent of reaction. A process flow diagram is shown for an SCR system in Figure 1.2.1-1. Flue gas is taken from the boiler between the economizer and air preheater. Ammonia, taken from a liquid storage tank and vaporized, is injected and mixed with the flue gas prior to the reactor. The flue gas passes through the catalyst bed where $NO_x$ is reduced to $N_2$. The flue gas then exits the reactor and is sent to the air preheater and, if necessary, further treatment equipment.

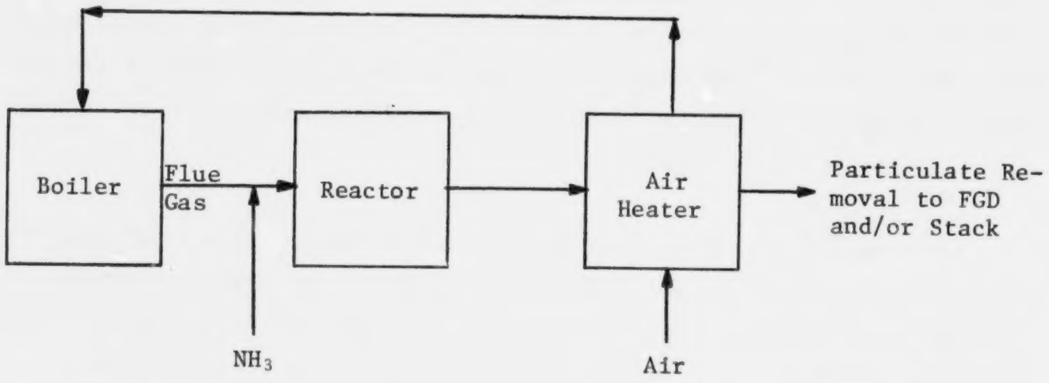

Figure 1.2.1-1. Flow diagram for typical $NO_x$-only SCR process.[5]

With this and all SCR systems it is desirable to treat flue gas exiting the economizer at 350-400°C prior to any air preheater since it is at this temperature range than the catalysts show the optimum combination of activity and selectivity. The analyses conducted in this study assumed that the boilers were operated constantly at full load and, therefore, had constant flue gas temperatures. However, it is possible that the boiler may experience large and frequent load swings which result in a variable flue gas temperature. FGT systems in this service will require flue gas heating in order to maintain sufficiently high temperatures. Temperature control can be accomplished by either a heater or a slipstream around the economizer. The heater will effectively decouple the FGT system from the boiler and does not require flow control of a flue gas slipstream. The economizer bypass will not derate the boiler since it will only be required during low load situations. In each of these approaches, much of the heat added to the flue gas will be captured in the air preheater. Both alternatives do, however, present an additional economic impact.

SCR systems can generally be applied to all boiler sizes and types, although with existing boilers there may be problems with spacial limitations. All of the catalysts considered here for use in treating flue gas containing $SO_2$ and $SO_3$ are resistant to poisoning by these compounds. Long term tests of these catalysts in the presence of $SO_x$ have shown very little or no decrease in activity or selectivity. Reactor size is proportional to flue gas flow rate, and this will determine the size and cost of the SCR system while the particulate concentration will determine the necessary catalyst/reactor combination.

The reactor itself is usually sized on the basis of a space velocity which is defined as the gas flow rate divided by the catalyst volume. A typical space velocity for a parallel flow system is about 6000 $hr^{-1}$ compared to 8000 $hr^{-1}$ for a moving bed or fixed, packed bed SCR system. The pressure drop through parallel flow systems is typically on the order of 100 mm $H_2O$, which is somewhat higher than moving or fixed, packed bed systems. The pressure drop is being reduced as this technology develops.

Parallel flow, moving bed and fixed, packed bed SCR systems are all capable of attaining the stringent level of $NO_x$ control. Greater than 90 percent $NO_x$ reduction is achieved at $NH_3$:$NO_x$ mole ratios of 1:1 on commercial systems applied to industrial boilers in Japan. All of these systems have been applied to a variety of oil-fired industrial boilers in Japan and appear to be viable techniques of attaining up to 90 percent $NO_x$ control.

There are some potential problems downstream of the FGT systems due to the presence of the unreacted ammonia in the flue gas. Two things can happen: 1) the $NH_3$ can react with $SO_2$ or $SO_3$ to form ammonium bisulfate or ammonium sulfate or 2) the $NH_3$ can enter the downstream equipment unreacted. The bisulfate has been shown to cause air preheater pluggage and this is the subject of ongoing research both at the EPA and the Electric Power Research Institute (EPRI). Both the bisulfate and sulfate exist as a particulate, but may be difficult to collect if the particles are submicron in size. Unreacted $NH_3$ is not likely to present any operational problems. A recent study has shown that if an ESP exists downstream, then most of the $NH_3$ will exit with the ash. $NH_3$ can actually improve the performance of an FGD system.[16]

### 1.2.2 Economic Impacts

The costs of $NO_x$ FGT systems applied to the industrial boilers are presented in this section. Two types of data are presented. First the capital and annual costs are shown as a function of boiler size. Then the cost effectiveness in terms of \$/kg $NO_x$ removed is evaluated. Tables 1.2.2-1 through 1.2.2-3 show the range of annual cost for the moderate to stringent level of control for the various boiler/size/control system combinations.

The annual costs in terms of \$/MBtu/hr are plotted against boiler size in Figures 1.2.2-1 through 1.2.2-4. In all cases, there is clearly an economy of scale with the larger units. An interesting result is that for the small residual oil-fired boiler, the parallel flow system is somewhat less expensive, but with the larger boiler, the moving bed system is less expensive. This is a result of the labor cost, which is a fixed cost, and is higher for moving bed systems than for parallel flow systems. Therefore, with small systems, the labor component has a significant effect on the annual cost of these systems. This result is the primary reason why it is not possible to choose a best system for residual oil applications. The effect of fuel type on annual cost is shown in Figure 1.2.2-5 when costs for the 44 $MW_t$ (150 MBtu/hr) boilers are compared for each fuel type. Systems applied to coal-fired boilers are the most expensive while those

TABLE 1.2.2-1. ANNUAL COST OF $NO_x$ CONTROL SYSTEMS APPLIED TO COAL-FIRED BOILERS

| Boiler | Size, MBtu/hr | Control System | Annual Cost, $1000/yr Moderate | Annual Cost, $1000/yr Stringent |
|---|---|---|---|---|
| Underfeed Stoker | 30 | Parallel Flow SCR | 108 | 130 |
| Chaingrate | 75 | Parallel Flow SCR | 153 | 197 |
| Spreader Stoker | 150 | Parallel Flow SCR | 221 | 291 |
| Pulverized Coal | 200 | Parallel Flow SCR | 254 | 351 |

TABLE 1.2.2-2. ANNUAL COST OF $NO_x$ CONTROL SYSTEMS APPLIED TO OIL FIRED-BOILERS

| Boiler | Size, MBtu/hr | Control System | Annual Cost, $1000/yr Moderate | Annual Cost, $1000/yr Stringent |
|---|---|---|---|---|
| Distillate Oil | 15 | Fixed Packed Bed SCR | 64 | 67 |
| Distillate Oil | 150 | Fixed Packed Bed SCR | 137 | 176 |
| Residual Oil | 30 | Parallel Flow SCR | 96 | 108 |
| Residual Oil | 30 | Moving Bed SCR | 120 | 130 |
| Residual Oil | 150 | Parallel Flow SCR | 181 | 223 |
| Residual Oil | 150 | Moving Bed SCR | 168 | 204 |

TABLE 1.2.2-3. ANNUAL COST OF $NO_x$ CONTROL SYSTEMS APPLIED TO NATURAL GAS-FIRED BOILERS

| Boiler | Size, MBtu/hr | Control System | Annual Cost, $1000/yr Moderate | Annual Cost, $1000/yr Stringent |
|---|---|---|---|---|
| Package, Firetube | 15 | Fixed Packed Bed SCR | 64.4 | 67.6 |
| Package, Watertube | 150 | Fixed Packed Bed SCR | 129 | 175 |

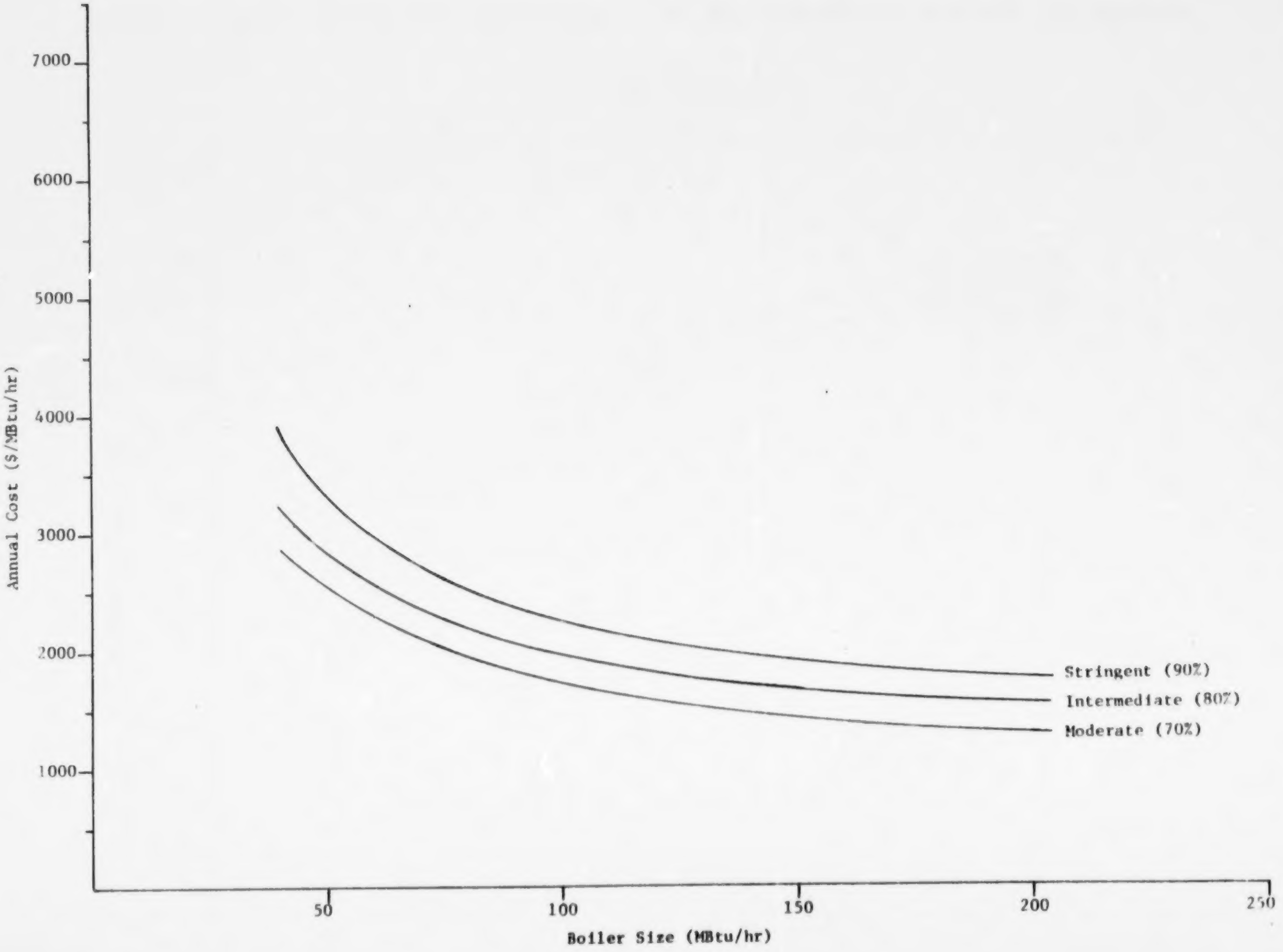

Figure 1.2.2-1. Annual cost of parallel flow SCR $NO_x$ FGT systems for coal-fired boilers.

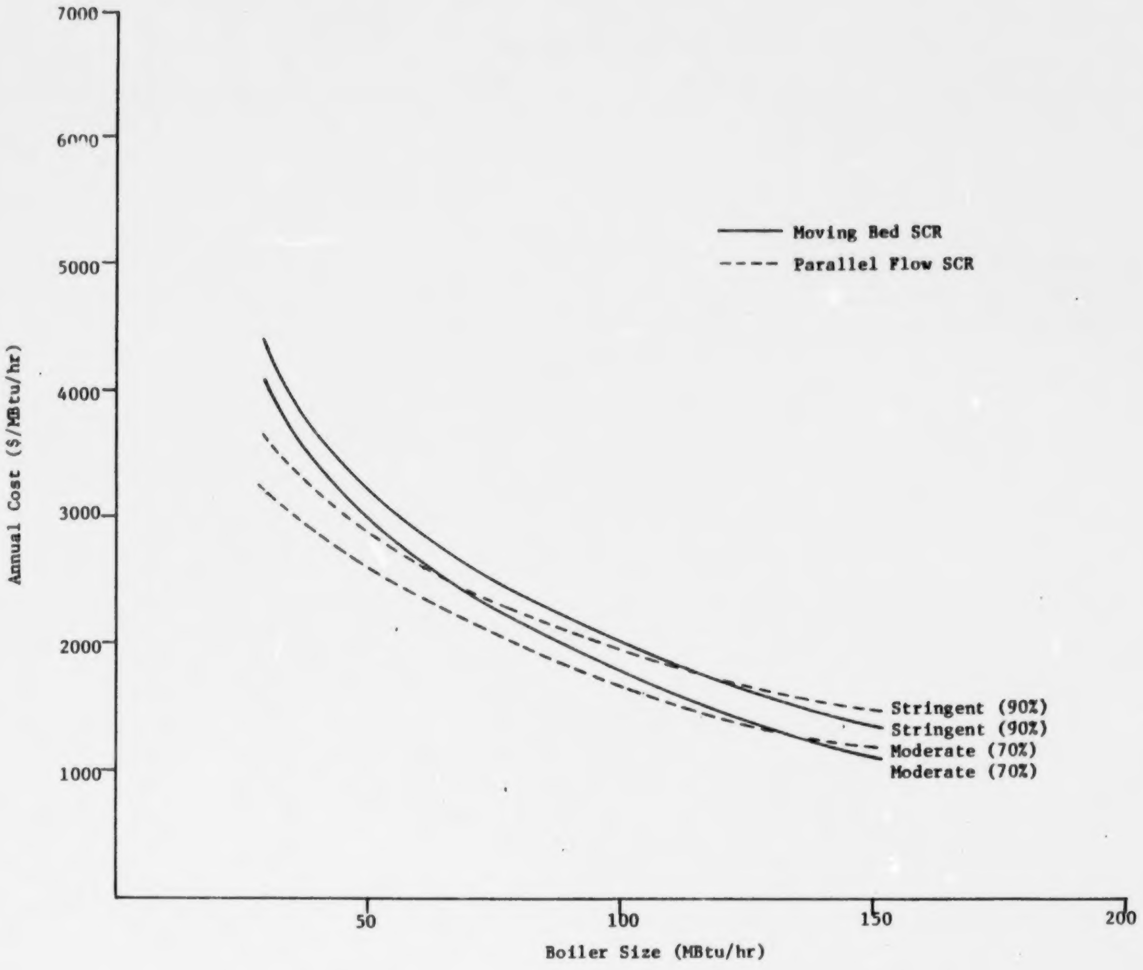

Figure 1.2.2-2. Annual cost comparison of $NO_x$ FGT systems for residual oil boilers.

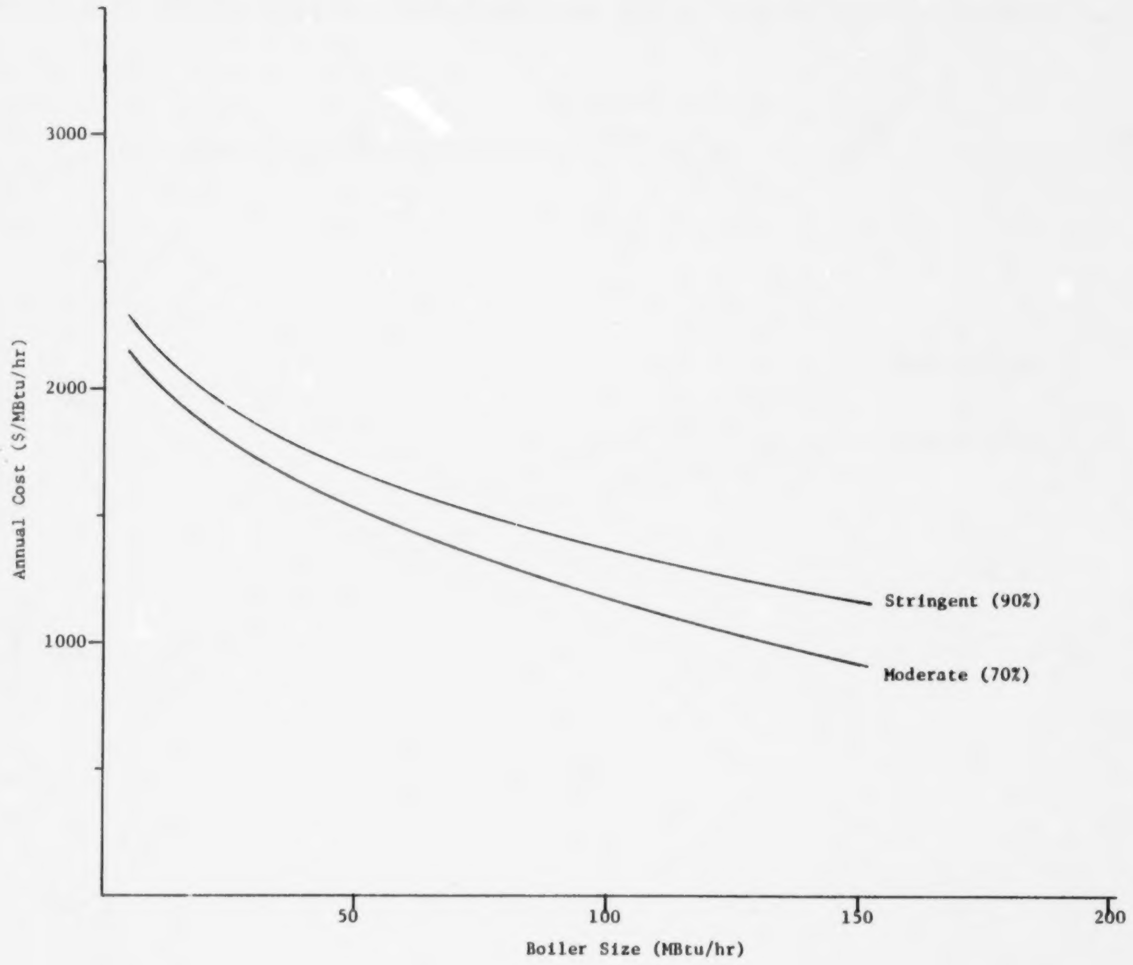

Figure 1.2.2-3. Annual cost of fixed packed bed SCR $NO_x$ FGT systems for distillate oil boilers.

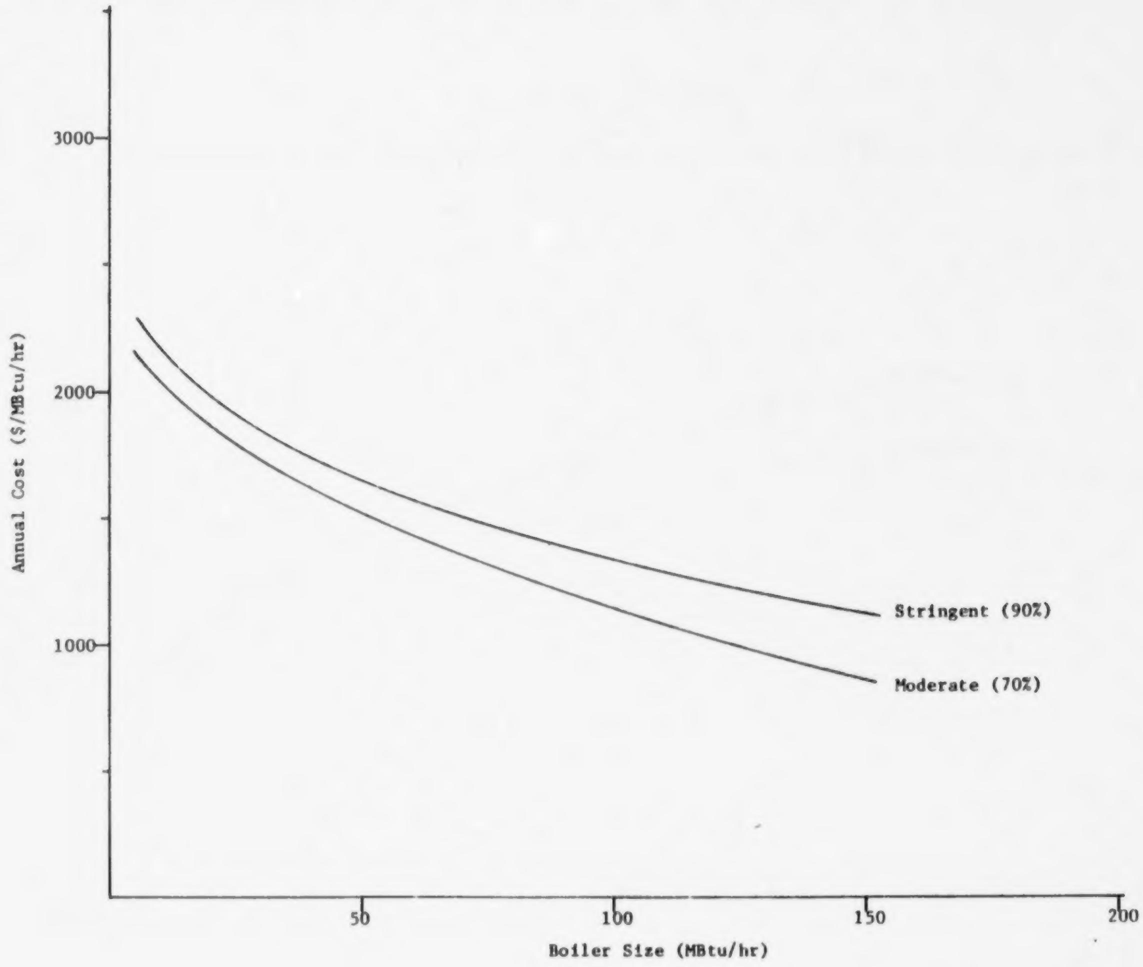

Figure 1.2.2-4. Annual cost of fixed packed bed SCR $NO_x$ FGT systems for natural gas boilers.

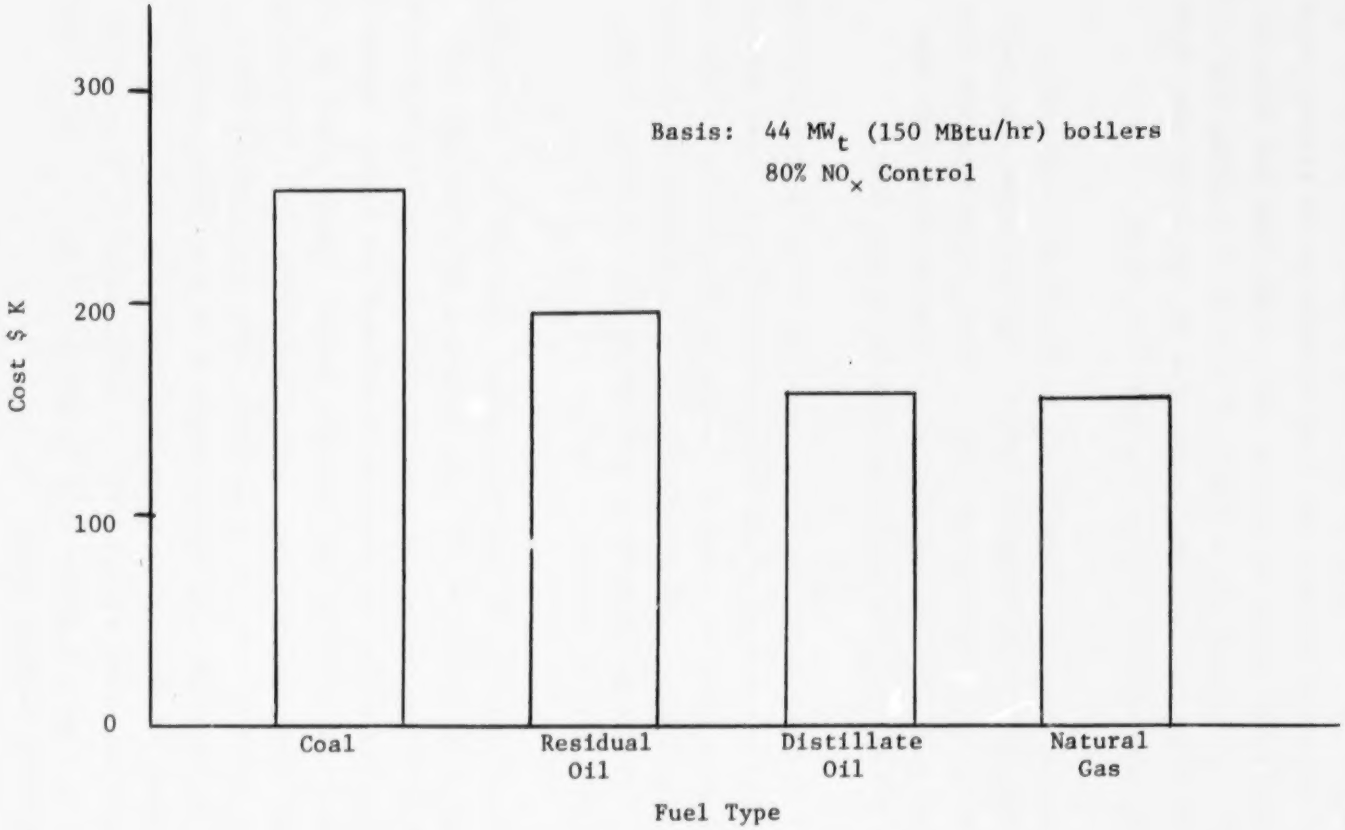

Figure 1.2.2-5. Comparison of annual costs of $NO_x$ FGT systems applied to 150 MBtu/hr boilers. (Average costs used where two systems apply.)

applied to distillate oil- and natural gas-fired boilers are the least expensive. Annual costs for residual oil-fired boilers lie in between these two extremes. The higher costs for systems which treat flue gas from coal- and residual oil-fired boilers are a result of the use of systems that will handle high particulate loadings, higher baseline $NO_x$ emissions with these fuels and, in the case of coal, higher flue gas flow rates.

Capital costs in terms of \$/MBtu/hr are presented as a function of boiler size in Figures 1.2.2-6 through 1.2.2-9. These figures also show larger systems to be less expensive in terms of cost per unit of capacity. This is due to the fact that the equipment costs used in this study were either constant for all sizes or varied exponentially with size.

In addition to determining the annual and capital costs, the study also examines the cost effectiveness of the various combinations. Cost effectiveness is defined as \$/kg $NO_x$ removed. Comparing the systems in this manner shows which combinations provide the largest environmental benefit for the lowest cost. Cost effectiveness if plotted against the level of $NO_x$ control in Figures 1.2.2-10 through 1.2.2-13.

For coal-fired boilers, the increased annual cost over an uncontrolled boiler for $NO_x$-only parallel flow SCR systems ranges approximately 6-12 percent, depending on the boiler and level of control. The figure plainly shows economy of scale as the largest coal-fired standard boiler, pulverized coal, has the most cost effective $NO_x$ control system. Annual costs for the small boilers are labor cost-dominant, hence the maximum cost effectiveness at 90 percent $NO_x$ control. The large boiler's costs are catalyst cost-dominant, hence the maximum cost effectiveness at 70 percent $NO_x$ control (additional catalyst is required to remove the additional $NO_x$). Similar effects occur with the other fuels as well. In all cases it is apparent that the system size has a significantly larger effect on the cost effectiveness than does the control level.

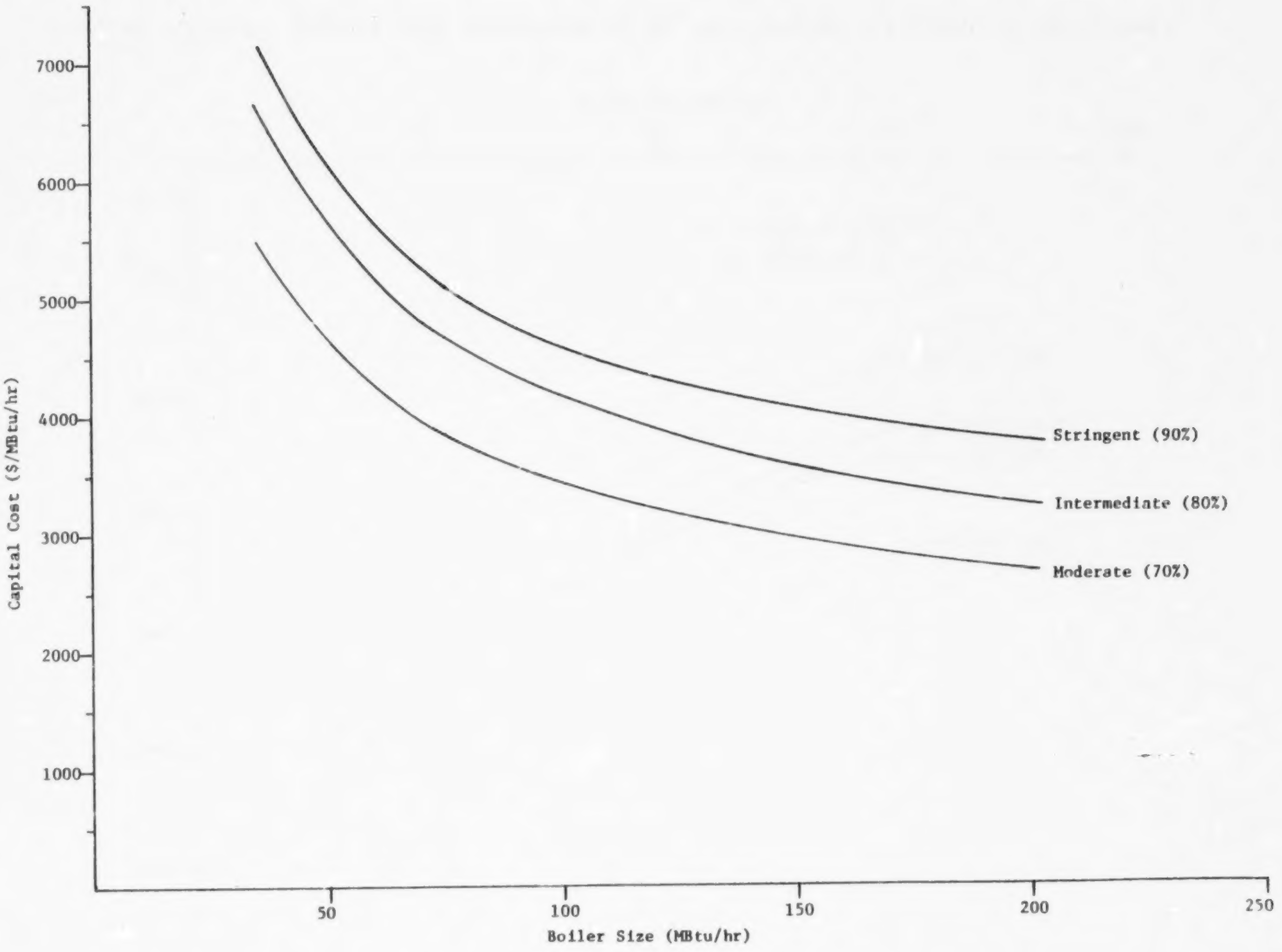

Figure 1.2.2-6. Capital cost of parallel flow SCR $NO_x$ FGT for coal-fired boilers.

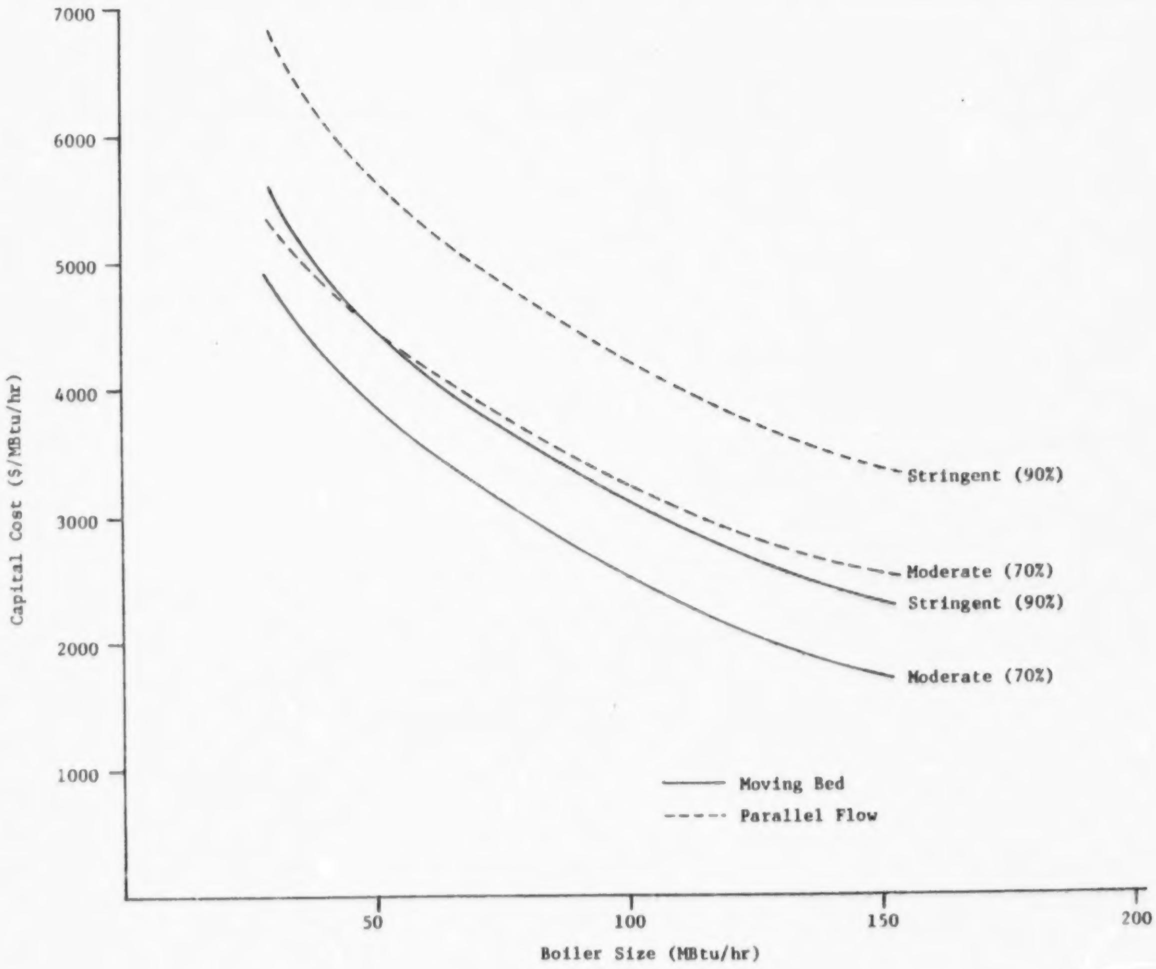

Figure 1.2.2-7. Capital cost comparison of $NO_x$ FGT systems for residual oil boilers.

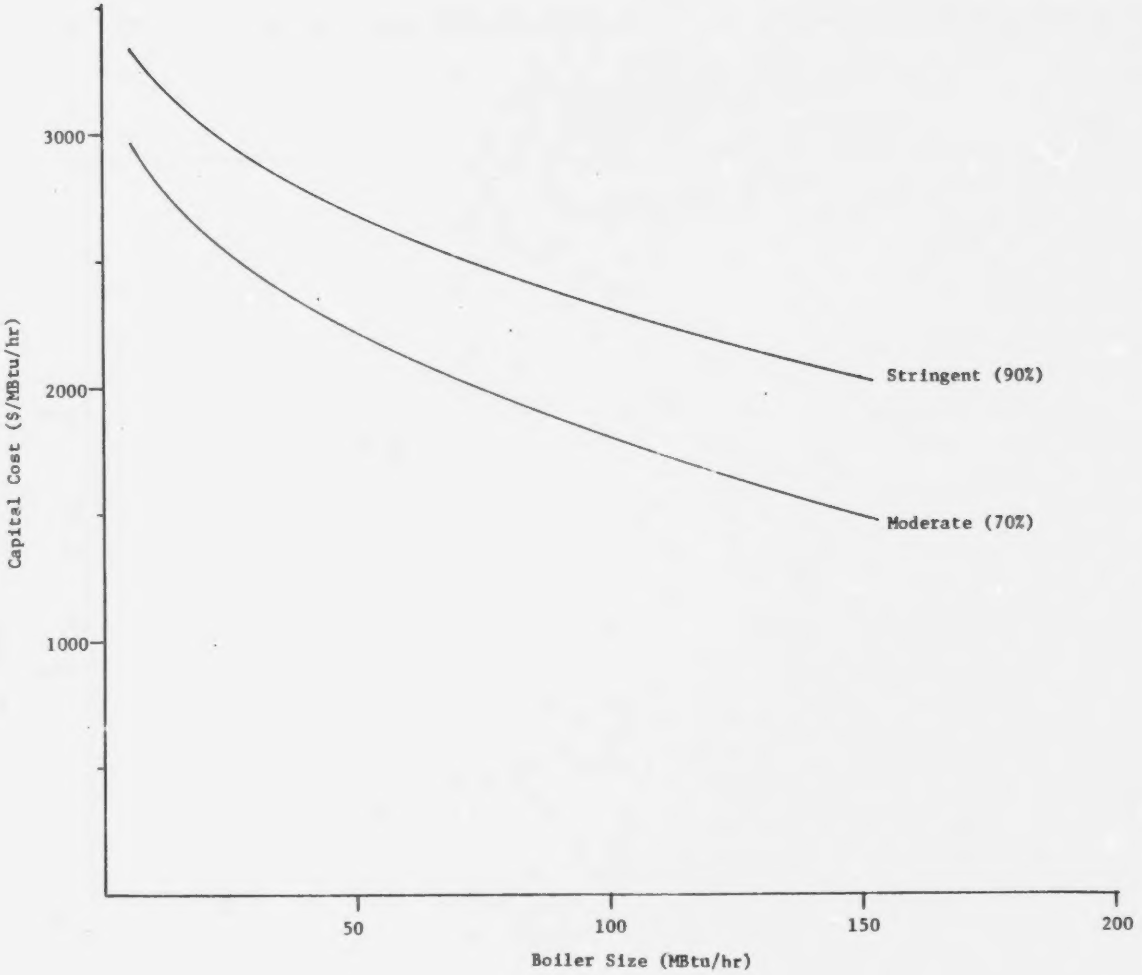

Figure 1.2.2-8. Capital cost of fixed packed bed SCR $NO_x$ FGT systems for distillate oil boilers.

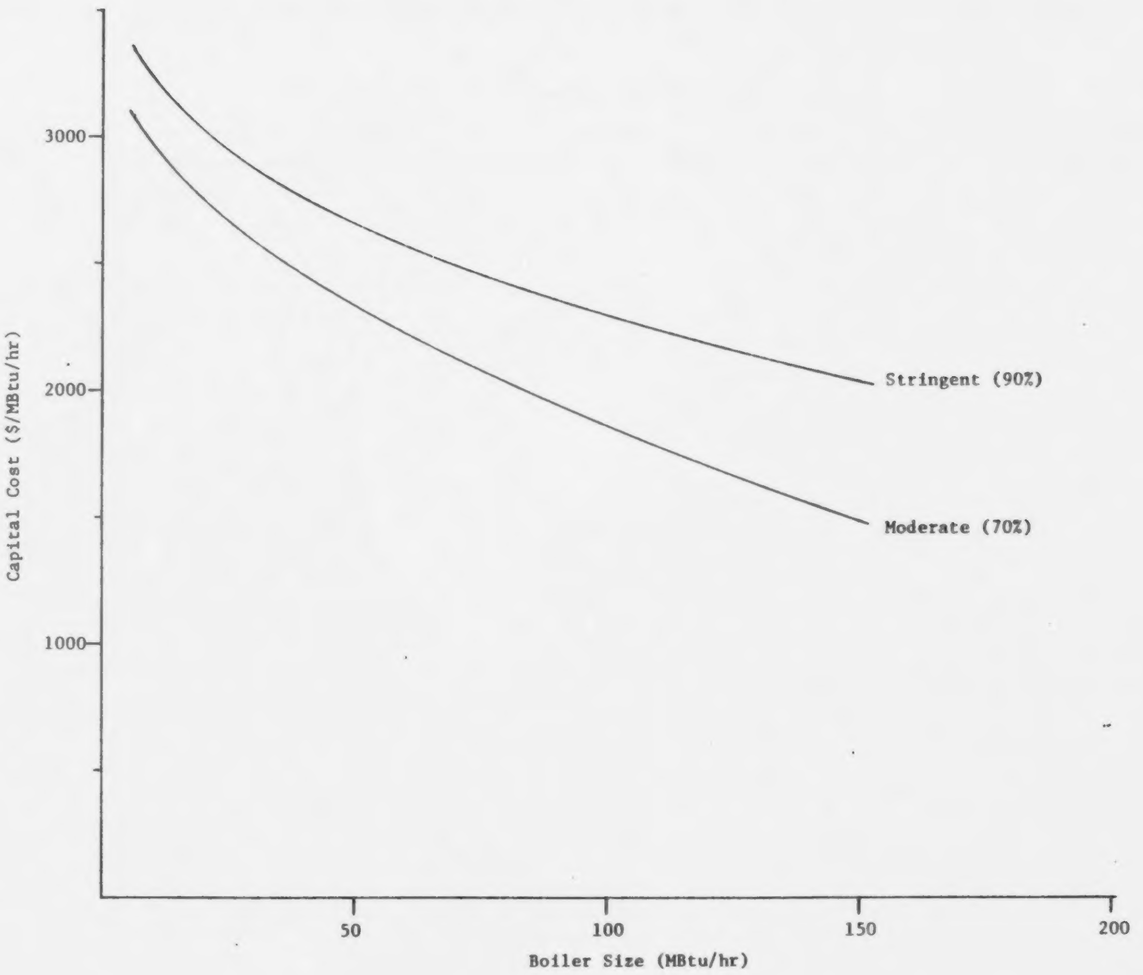

Figure 1.2.2-9. Capital cost of fixed packed bed SCR $NO_x$ FGT systems for natural gas boilers.

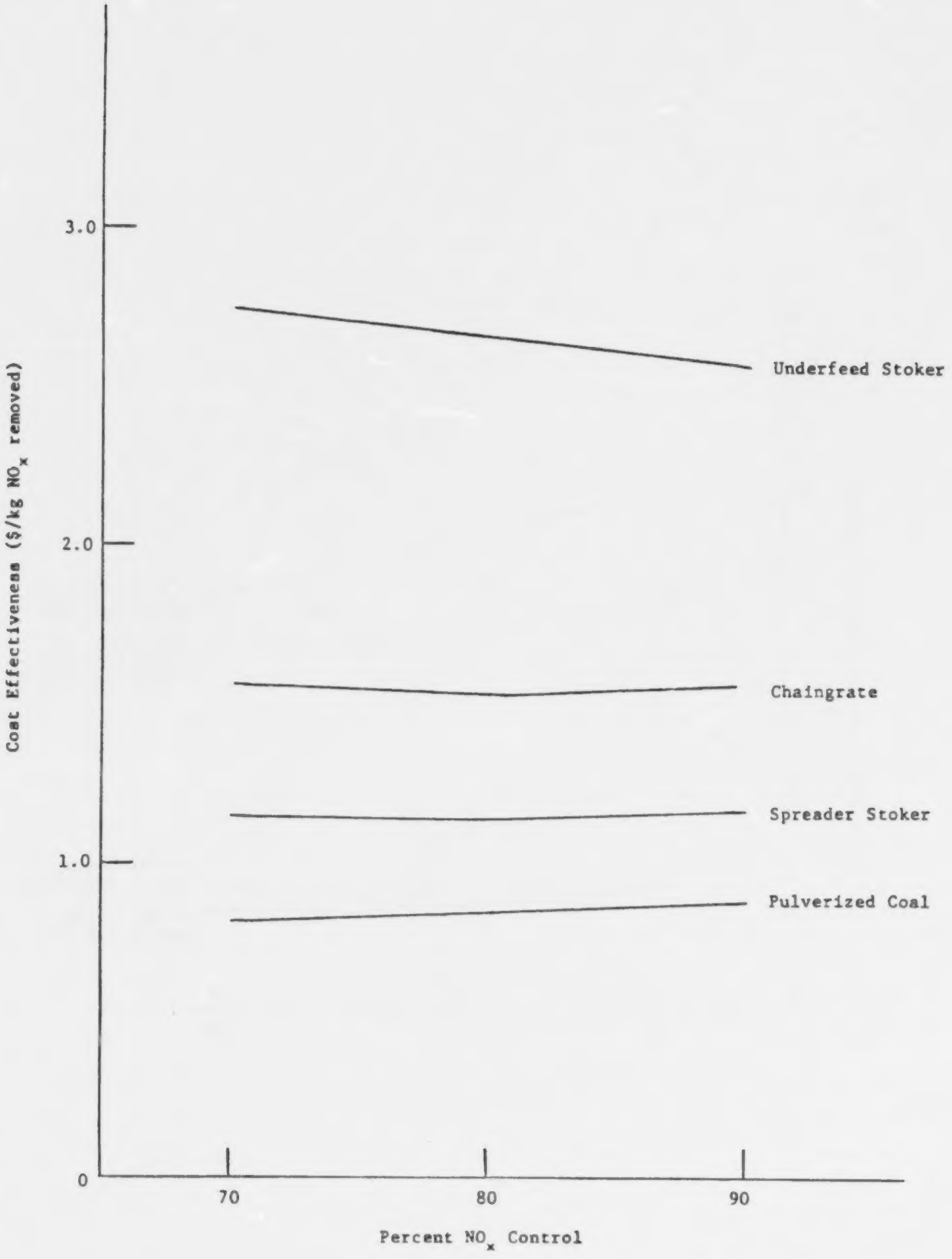

Figure 1.2.2-10. Cost effectiveness of parallel flow NO$_x$ control systems for coal-fired boilers.

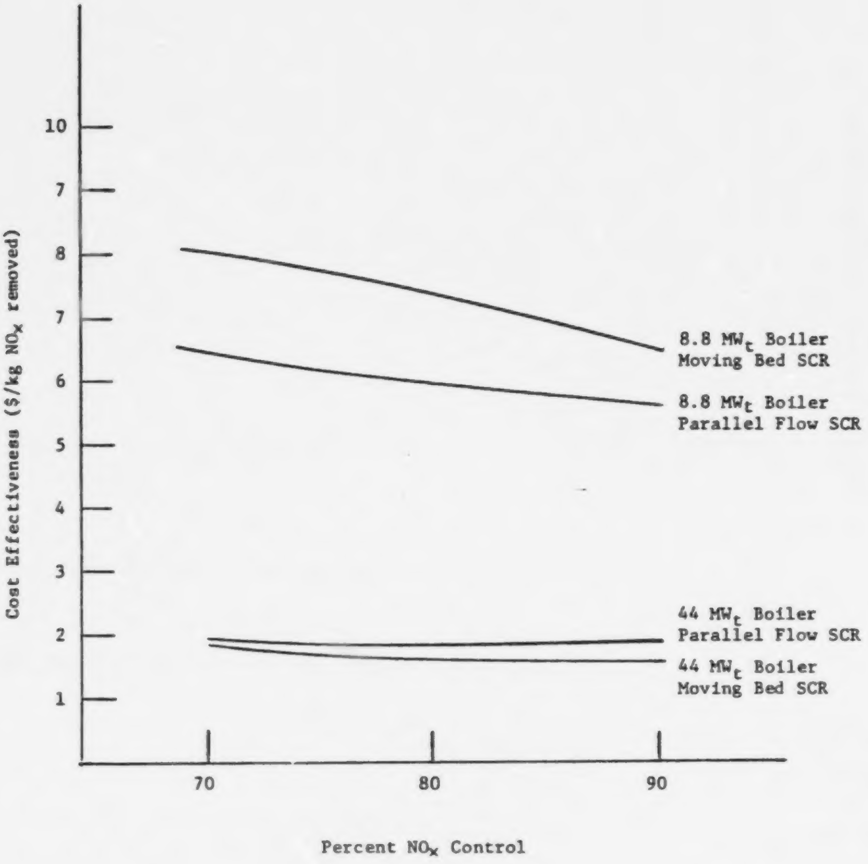

Figure 1.2.2-11. Cost effectiveness of FGT systems applied to residual oil-fired boilers.

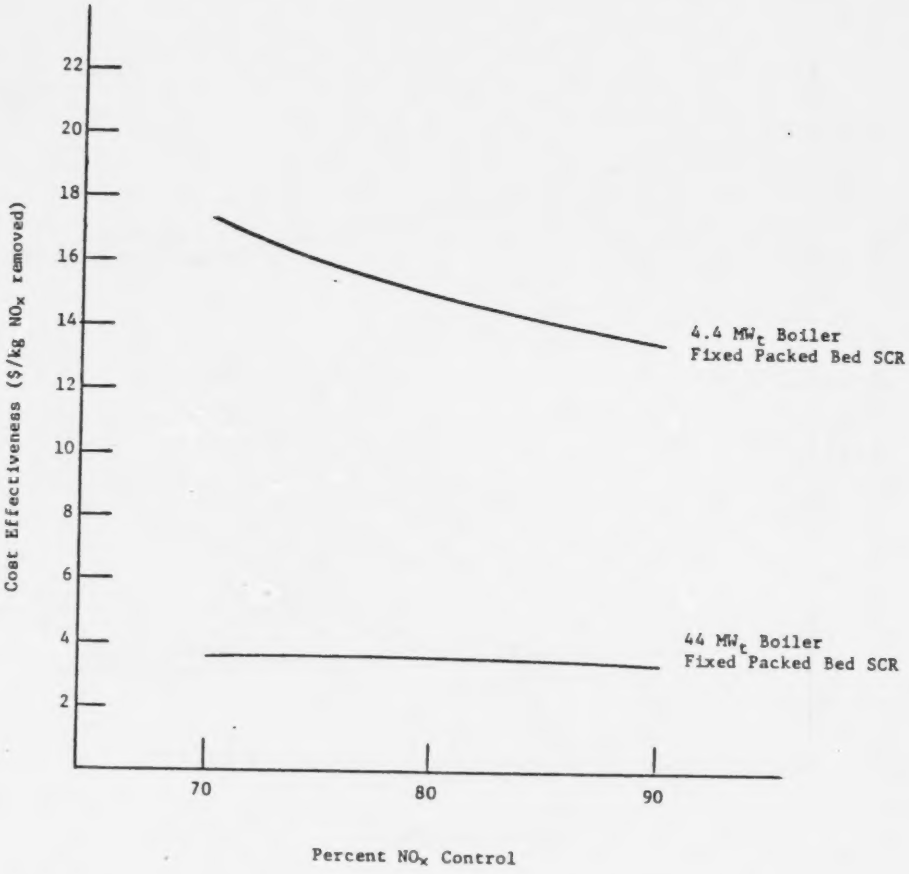

Figure 1.2.2-12. Cost effectiveness of FGT systems applied to distillate oil-fired boilers.

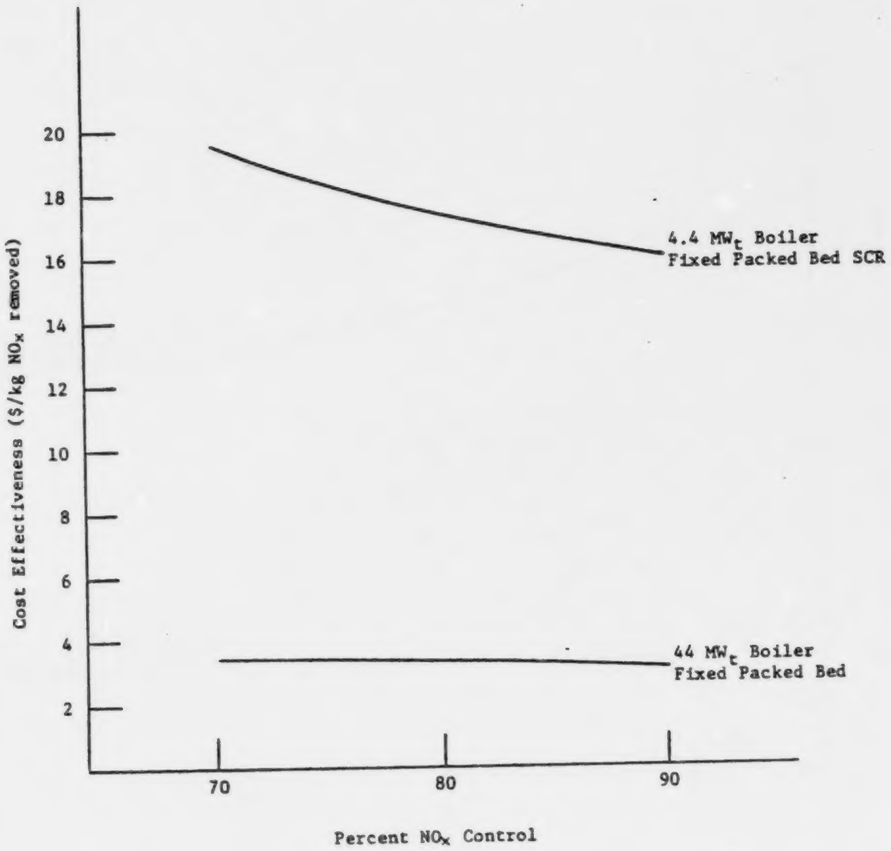

Figure 1.2.2-13. Cost effectiveness of FGT systems applied to natural gas-fired boilers.

The costs of SCR applications to modified or reconstructed facilities will be higher than those shown here. It is estimated that these costs will range from 25 to 120 percent more than applications to new boilers.

### 1.2.3 Energy Impacts

In calculating energy usage for each of the cases, all sources of energy consumption were considered. These sources are shown in Table 1.2.3-1.

TABLE 1.2.3-1. AREAS OF ENERGY CONSUMPTION IN $NO_x$ FGT SYSTEMS

| $NO_x$ FGT System | Energy Consumption Step (equipment) | Type of Energy Consumed |
|---|---|---|
| Parallel Flow SCR | Reactor Draft Loss (Fan) | Electrical |
| | Liquid $NH_3$ Transfer (Pump) | Electrical |
| | $NH_3$ Vaporization (Vaporizer) | Steam |
| | $NH_3$ Dilution | Steam |
| Moving Bed SCR | Reactor Draft Loss (Fan) | Electrical |
| | Liquid $NH_3$ Transfer (Pump) | Electrical |
| | Catalyst Screening & Transfer (Elevator) | Electrical |
| | Baghouse Draft Loss (Blower) | Electrical |
| | $NH_3$ Vaporization (Vaporizer) | Steam |
| | $NH_3$ Dilution | Steam |
| Fixed Packed Bed SCR | Reactor Draft (Fan) | Electrical |
| | Liquid $NH_3$ Transfer (Pump) | Electrical |
| | $NH_3$ Vaporization (Vaporizer) | Steam |
| | $NH_3$ Dilution | Steam |
| | Soot Blowing-Distillate Oil Boiler Only | Steam |

The energy impacts are presented in two forms. In the first, energy consumption in terms of MBtu/hr is plotted as a function of boiler size. These data are shown in Figures 1.2.3-1 through 1.2.3-4. Essentially, in all of the cases energy consumption is less than 1 MBtu/hr and represents a small amount of energy. The relative amount of energy consumed is shown in Figures 1.2.3-5 through 1.2.3-8 where usage is shown as a percent of the boiler heat input.

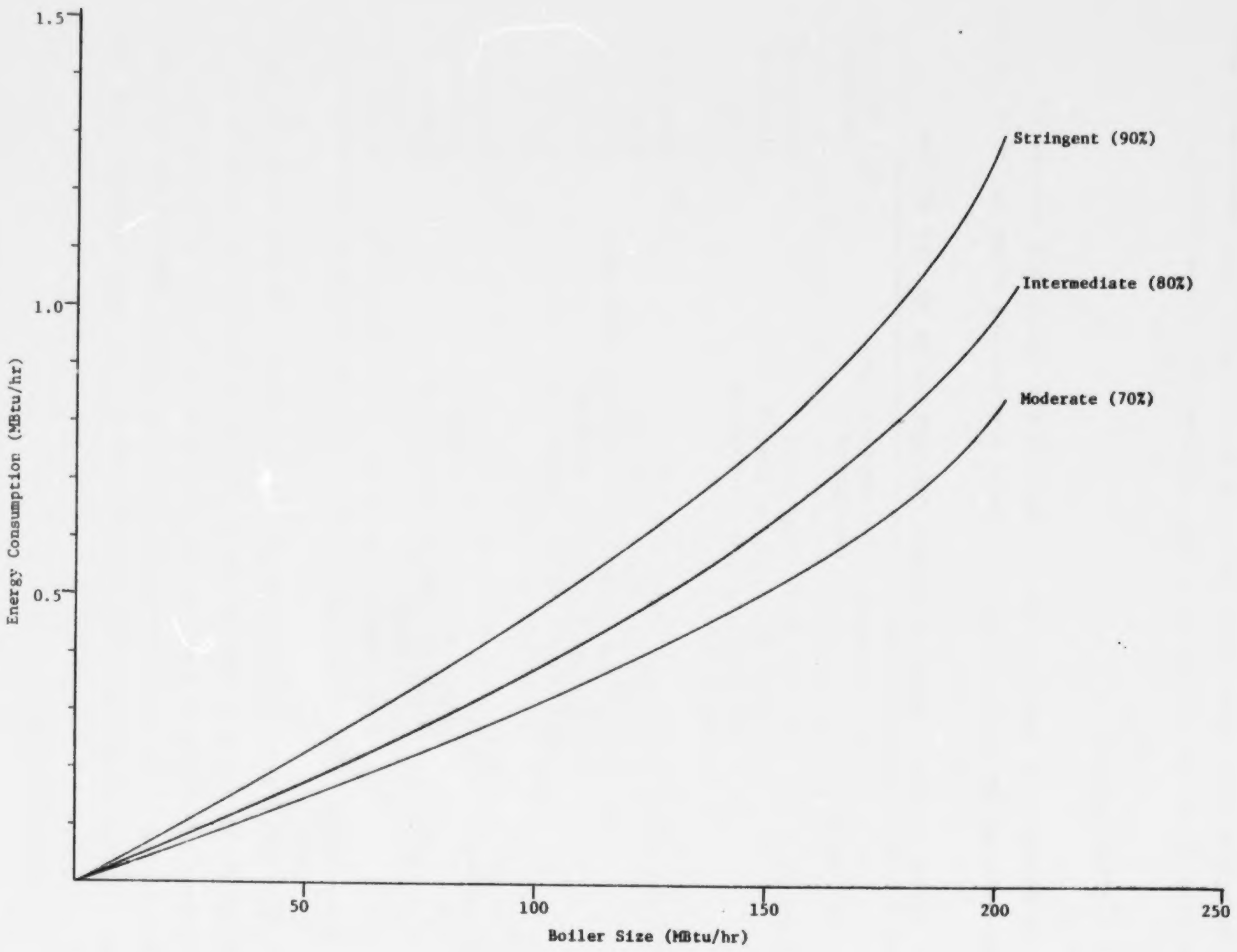

Figure 1.2.3-1. Energy consumption of parallel flow SCR $NO_x$ FGT systems for coal-fired boilers.

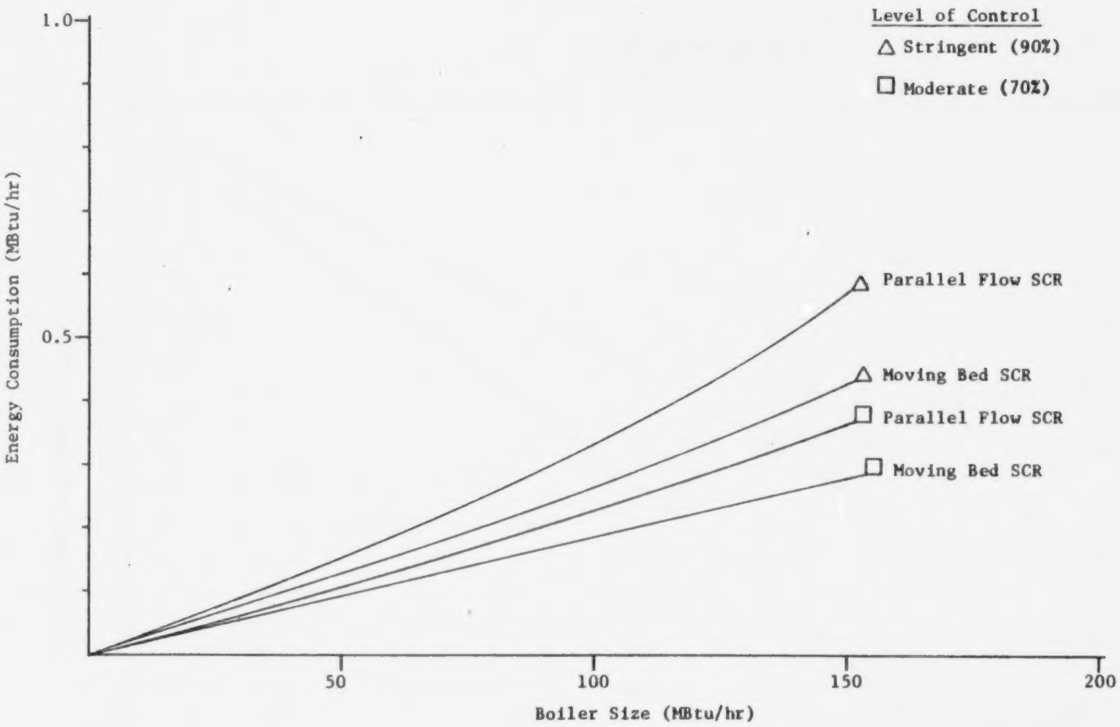

Figure 1.2.3-2. Energy consumption of $NO_x$ FGT systems for residual oil boilers.

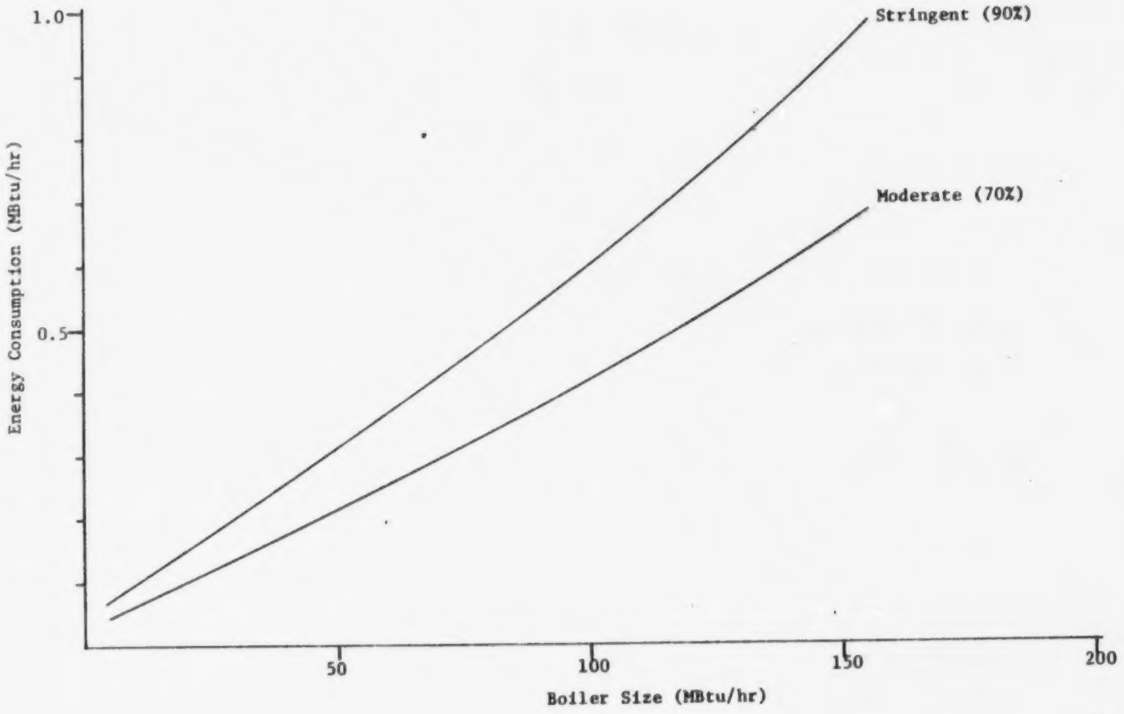

Figure 1.2.3-3. Energy consumption of fixed packed bed SCR $NO_x$ FGT systems for distillate oil boilers.

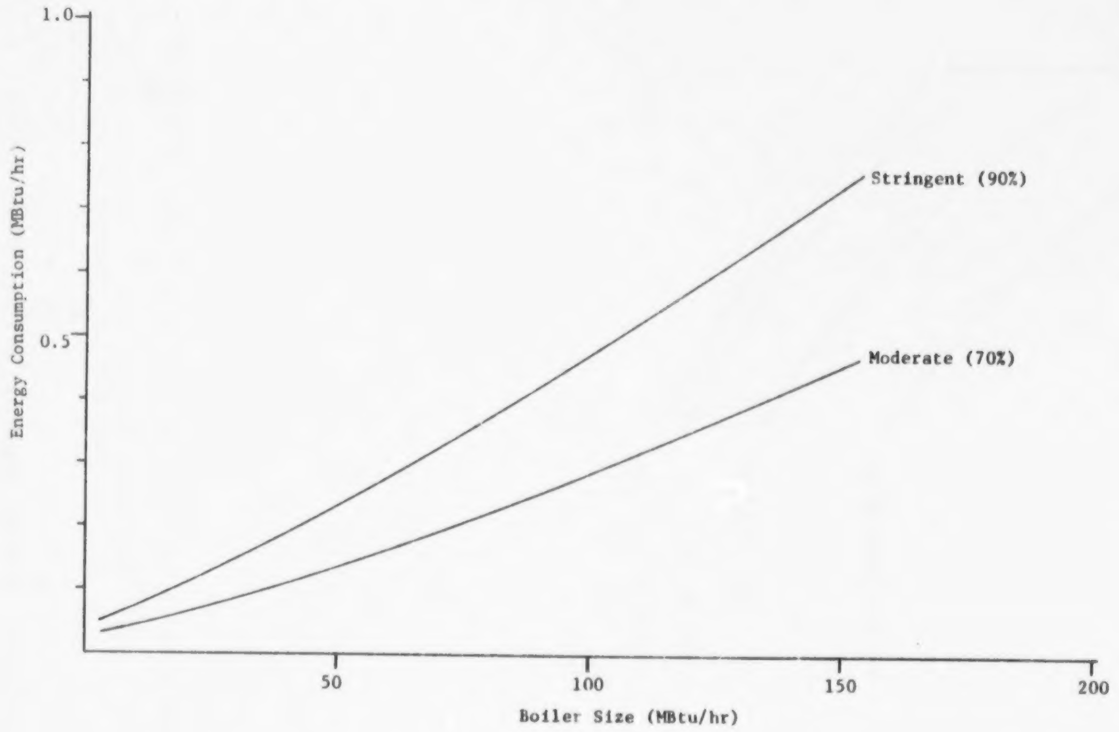

Figure 1.2.3-4. Energy consumption of fixed packed bed SCR $NO_x$ FGT systems for natural gas boilers.

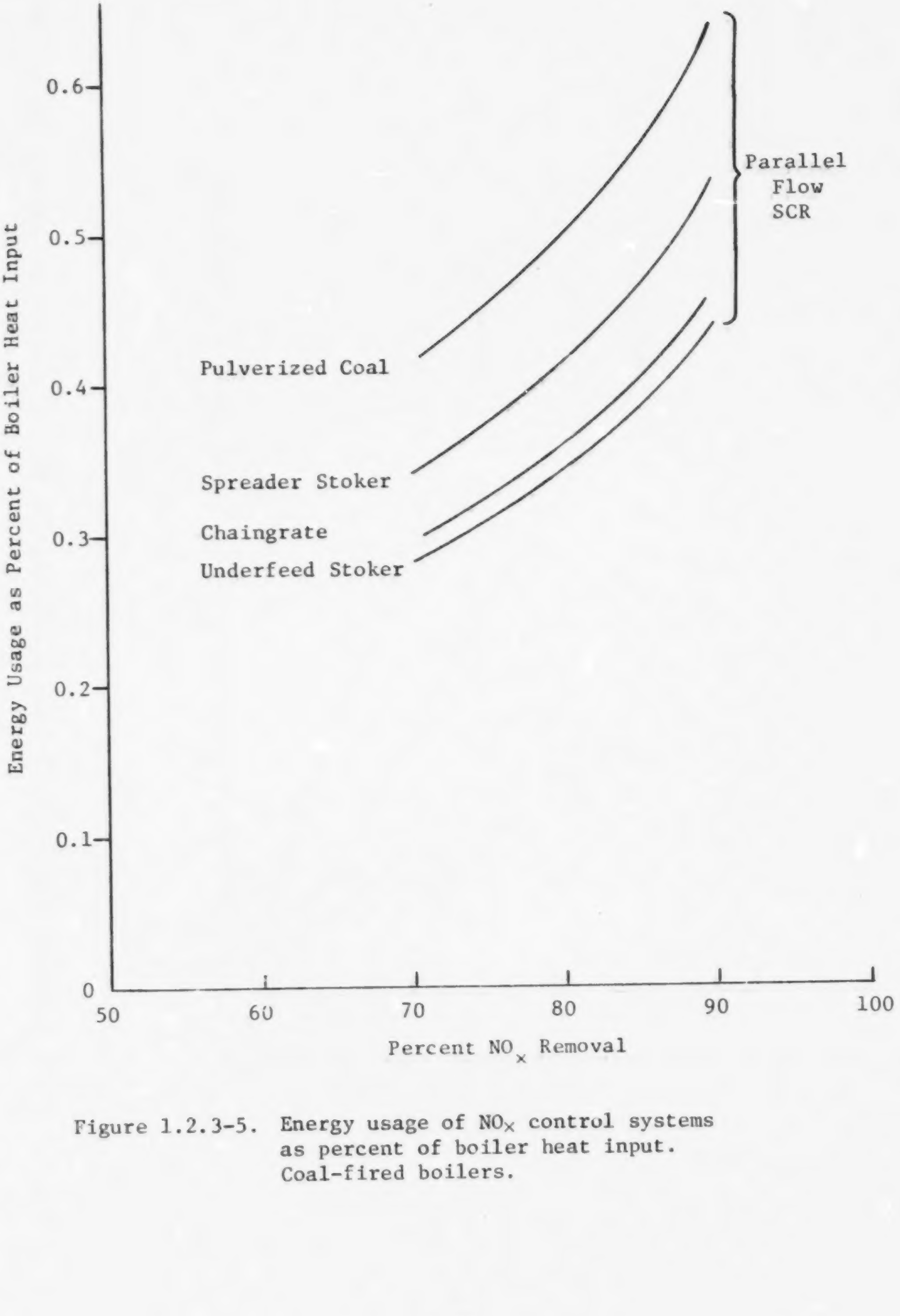

Figure 1.2.3-5. Energy usage of $NO_x$ control systems as percent of boiler heat input. Coal-fired boilers.

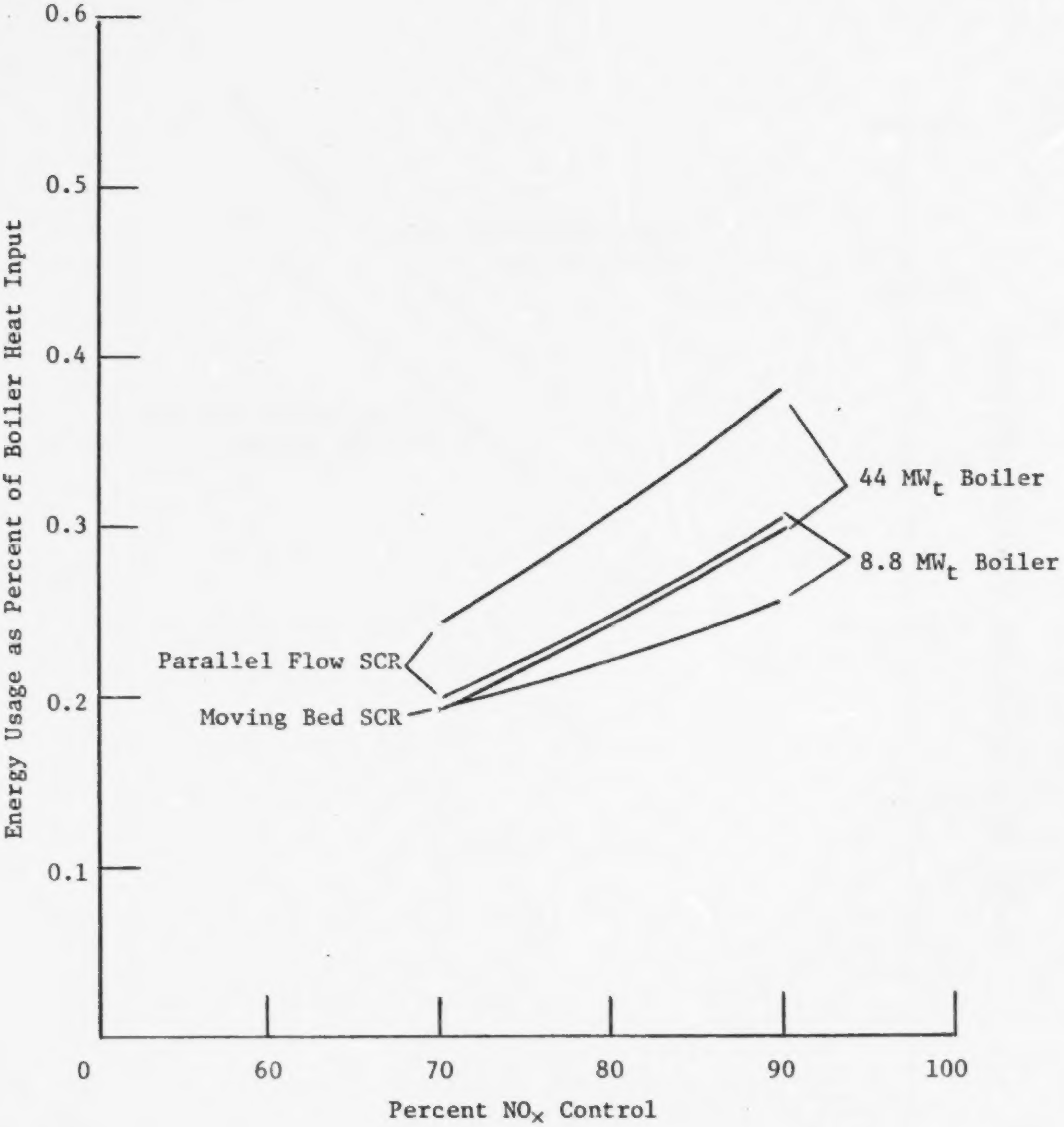

Figure 1.2.3-6. Energy usage of $NO_x$ control systems as percent of boiler heat input.
Residual oil-fired boilers.

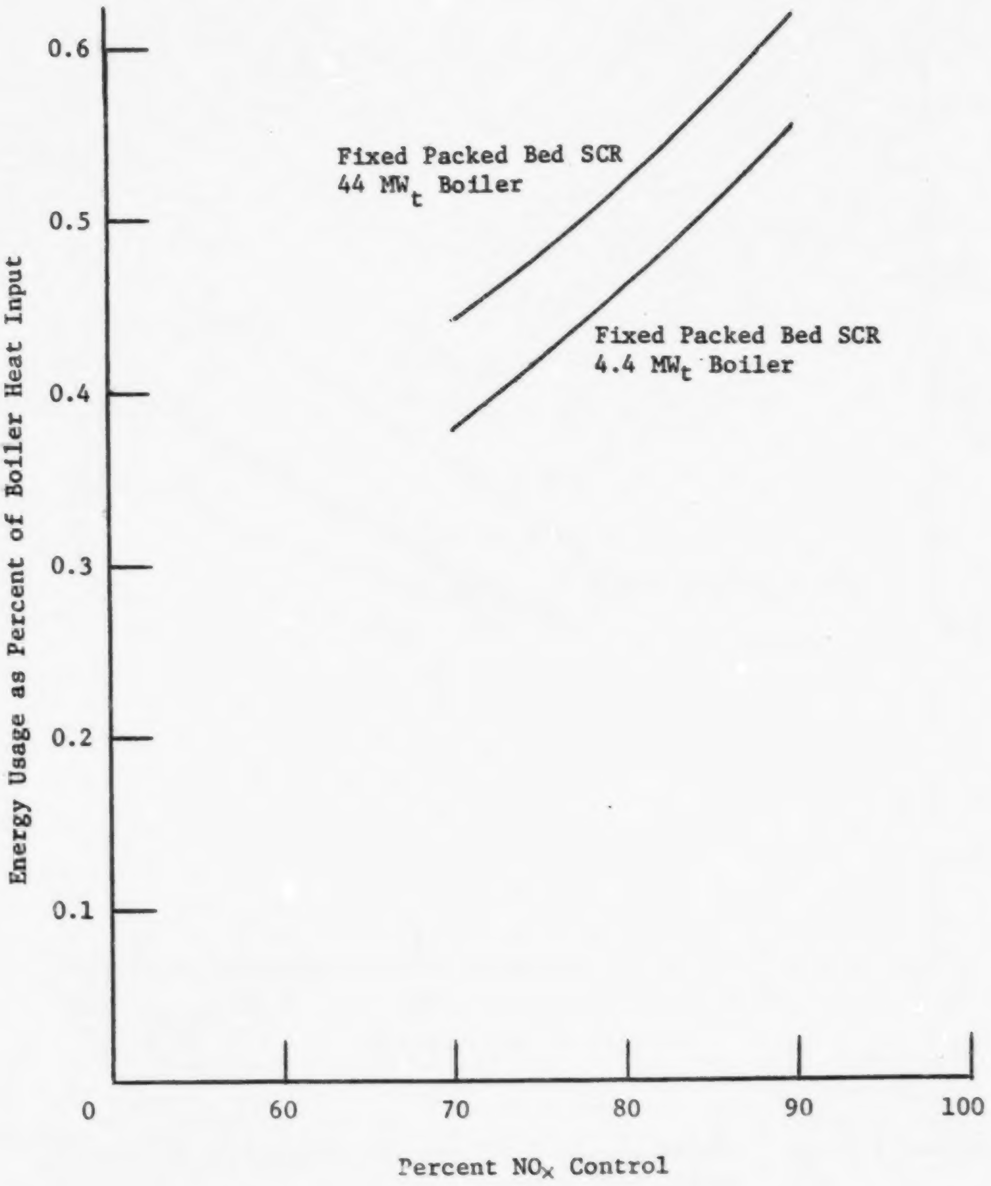

Figure 1.2.3-7. Energy usage of $NO_x$ control systems as a function of boiler heat input.
Distillate oil-fired boilers.

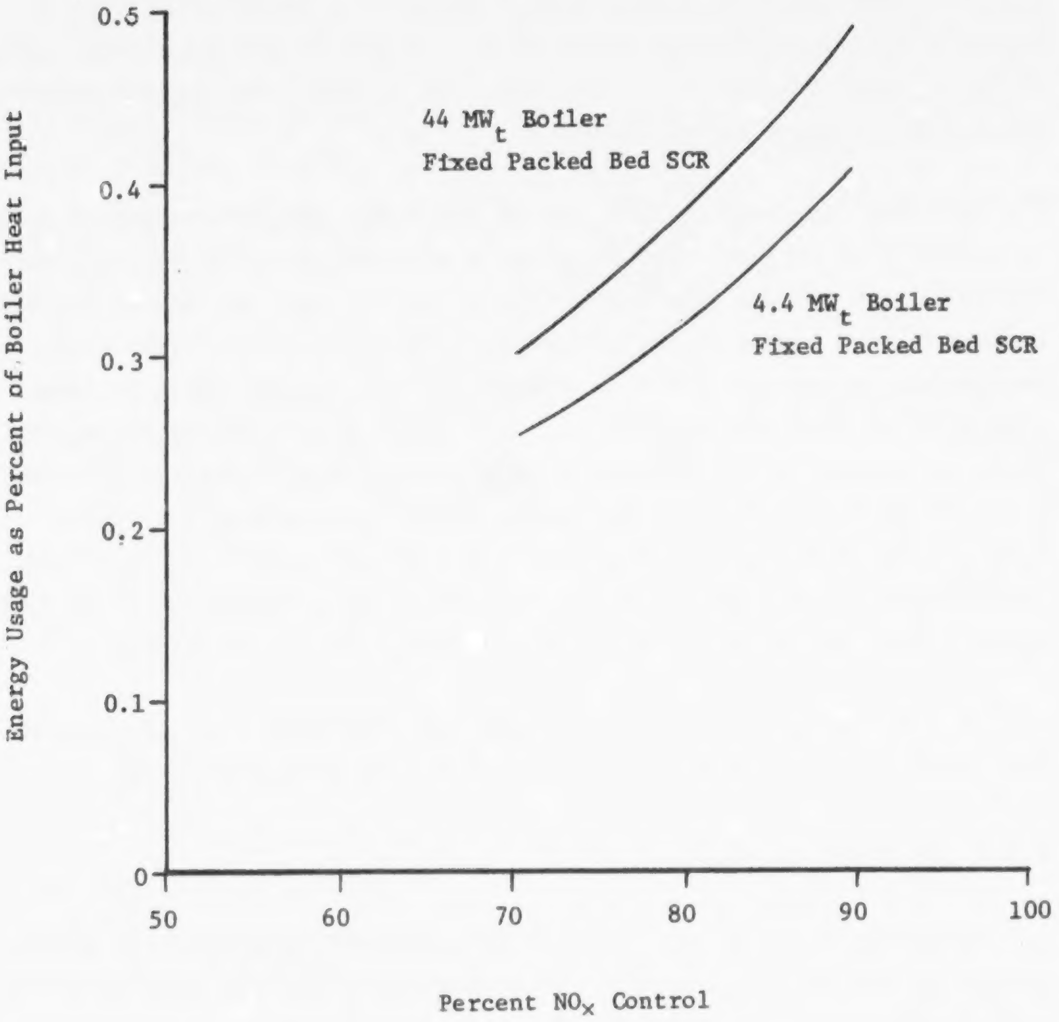

Figure 1.2.3-8. Energy usage of $NO_x$ control systems as percent of boiler heat input. Natural gas-fired boilers.

For the coal-fired boilers, the systems have a range of 0.27 to 0.64 percent. Energy usage is more a function of $NO_x$ control level than boiler size. In this analysis, energy usage is higher for larger boilers as a result of the reactor design method which allowed the reactor pressure drop to vary. Also, for the pulverized coal case, higher inlet $NO_x$ concentrations lead to higher energy usage.

With the oil-fired boilers, the parallel flow SCR systems have a range of 0.20 to 0.38 percent (from moderate to stringent level of control) and the moving bed systems from 0.19 to 0.29. One can see that the moving bed systems require less energy than the parallel flow systems. This is due to the greater pressure drop across a parallel flow reactor, which is larger than the $\Delta P$ across a moving bed reactor. Again, as in economics, the two candidate systems are considered to have similar energy impacts. For the distillate oil-fired boiler, the energy consumption ranged from 0.33 to 0.62 percent of the boiler heat input. For $NO_x$ FGT applied to gas-fired boilers, the small fixed packed bed SCR system has a range of 0.27 to 0.42 percent, whereas the large system varies from 0.30 to 0.49 percent.

In all cases, energy usage was less than 0.64 percent of the boiler heat input, and in most cases it was less than half of this amount.

### 1.2.4 Environmental Impacts

There are some potential adverse environmental impacts of SCR systems. The use of $NH_3$ as the gaseous reducing agent introduces the possibility of ammonia emissions. The level of $NH_3$ emissions experienced by commercial SCR operations range from 1 to 10 ppm depending on the control level. Even at elevated $NH_3$:$NO_x$ ratios (>1.0), the $NH_3$ emissions are reported to be less than 20 ppm. It is possible that $NH_3$ emissions will increase as the catalyst ages; however, commercial applications have not operated long enough to show this effect. Ten ppm of $NH_3$ may be an optimistic value, especially considering that currently there is no continuous monitoring technique for measuring

$NH_3$ in the presence of $SO_x$. The data, therefore, represent spot measurements and not continuous data. It seems reasonable to assume that 10 ppm represents a minimum level of $NH_3$ emissions. $NH_3$ emissions in terms of pounds per MBtu are presented in Figures 1.2.4-1 through 1.2.4-3 as a function of boiler size. While there is some variation, emission levels are essentially constant for all boiler sizes.

Another potential environmental problem is the formation of ammonium bisulfate, $NH_4HSO_4$, or ammonium sulfate, $(NH_4)_2SO_4$. The presence of $NH_3$, $SO_3$, and $H_2O$ in the hot flue gas leads to the formation of liquid $NH_4HSO_4$ upon cooling to approximately 180-220°C by the following reaction.

$$NH_3(g) + SO_3(g) + H_2O(g) \rightleftarrows NH_4HSO_4(l) \qquad (1\text{-}3)$$

This can create a plugging and corrosion problem in heat exchange equipment, particularly when medium- or high-sulfur fuels are fired. A beneficial effect is obtained by the tying up of $SO_3$ which is more hazardous than $SO_2$ and difficult to catch with FGD.[12] Further cooling to about 190°C precipitates the formation of solid ammonium sulfate by the following reaction.

$$NH_4HSO_4(l) + NH_3(g) \rightleftarrows (NH_4)_2SO_4(s) \qquad (1\text{-}4)$$

It is speculated that minor, if any, amounts of these sulfates will be emitted to the atmosphere in situations where particulate control equipment exists downstream of the $NO_x$ control system. Sulfate formation is not a problem with gas-fired boilers since there is no sulfur present in the fuel.

Disposal of spent catalyst is the final environmental concern of the parallel flow SCR systems. Catalysts such as titanium dioxide ($TiO_2$) and vanadium pentoxide ($V_2O_5$) are probably recycled due to their high cost. This question is currently unanswered since all applications of this technology are very recent and none have yet required a catalyst change.

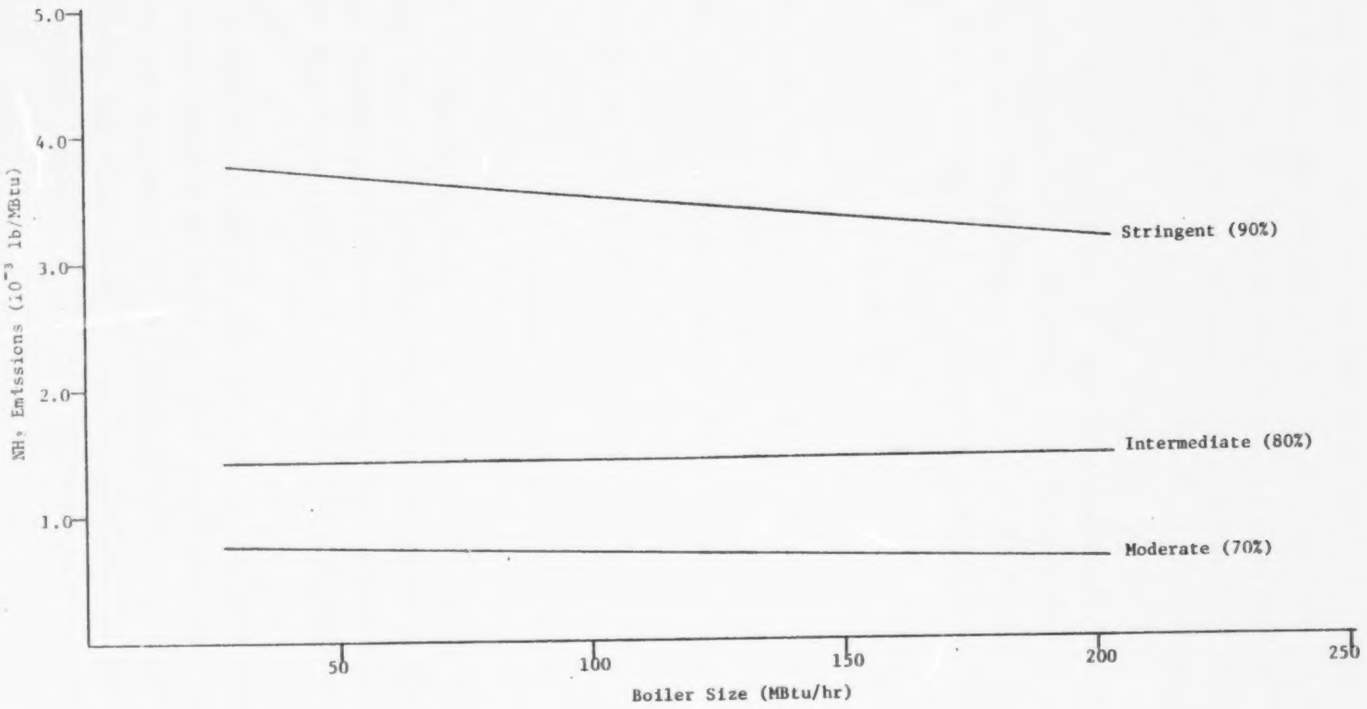

Figure 1.2.4-1. $NH_3$ emissions from SCR $NO_x$ FGT systems for coal-fired boilers.

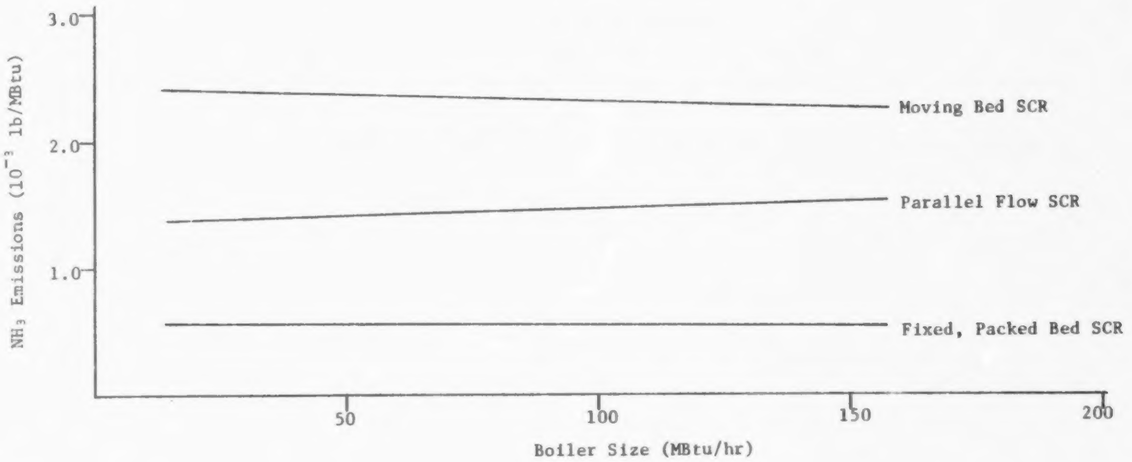

Figure 1.2.4-2. $NH_3$ emissions from SCR $NO_x$ FGT systems for oil-fired boilers.

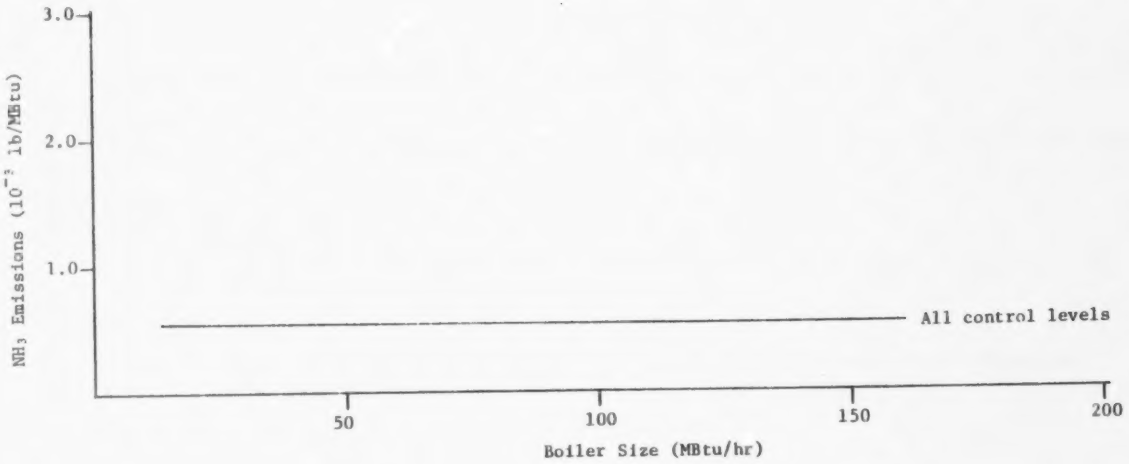

Figure 1.2.4-3. NH$_3$ emissions from SCR NO$_x$ FGT systems for natural gas-fired boilers.

Summarizing, FGT processes are relatively clean, possess minor potential air pollution and waste problems, and have no water, thermal, or noise pollution.

### 1.2.5 Development Status

Parallel flow SCR processes have been applied in Japan to several residual oil-fired industrial boilers. Oil-fired utility boilers and other sources with high particulate concentrations are also being treated. SCR processes have not yet been demonstrated commercially on coal-fired boilers. However, pilot units have been operated and some U.S. firms are offering SCR processes for use on coal-fired boilers. Two applications to coal-fired utility boilers are planned for 1980 (Table 1.2.5-1) although none exist at the present time. A coal-fired pilot unit demonstration of one parallel flow design is currently underway in the U.S. under EPA sponsorship[14] and several have been conducted in Japan. The EPA facility should be operational by early 1980. Another U.S. demonstration of a $NO_x$-only SCR process will be performed in 1980 by the Electric Power Research Institute.

TABLE 1.2.5-1. PLANNED FGT INSTALLATIONS OF SCR COAL-FIRED UTILITY BOILERS[13]

| Location | User | Process Developer | Fuel | Capacity ($Nm^3$/hr) | Completion Date |
|---|---|---|---|---|---|
| Takehara | Electric Power Development Co. | Has not been selected | Coal | 800,000 | July 1981 |
| Tomato | Hokkaido Electric | Hitachi, Ltd. | Coal | 280,000 | October 1980 |

Parallel flow and moving bed SCR processes have been applied in Japan to several oil-fired industrial and utility boilers. These operations are summarized in Tables 1.2.5-2 through 1.2.5-5. SCR systems are considered commercially available for oil-fired boilers at this time.

TABLE 1.2.5-2. EXISTING FGT INSTALLATIONS OF SCR PARALLEL FLOW SYSTEMS OIL-FIRED INDUSTRIAL BOILERS[17]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3$/hr) | Completion Date |
|---|---|---|---|---|---|
| Sodegaura | Fuji Oil | Mitsubishi H.I. | Resid | 200,000 | January 1978 |
| Kawasaki | Ajinomoto | Ishikawajima H.I. | Resid | 180,000 | January 1978 |
| Chiba | Ukishima Pet. Chem. | Mitsui Engineering | Resid | 220,000 | April 1978 |

TABLE 1.2.5-3. EXISTING FGT INSTALLATIONS OF SCR PARALLEL FLOW SYSTEMS OIL-FIRED UTILITY BOILERS[18]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3$/hr) | Completion Date |
|---|---|---|---|---|---|
| Yokosuka | Tokyo Electric | Mitsubishi H.I. | Resid | 40,000 | March 1977 |
| Chita | Chubu Electric | Mitsubishi H.I. | Resid | 1,920,000 | February 1980 |
| Kudamatsu | Chugoku Electric | Ishikawajima H.I. | Resid | 1,900,000 | July 1979 |
| Niigata | Tohoku Electric | Ishikawajima H.I. | Resid | 1,660,000 | August 1981 |

TABLE 1.2.5-4. EXISTING FGT INSTALLATIONS OF SCR MOVING BED SYSTEMS OIL-FIRED INDUSTRIAL BOILERS[18]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Kaizuka | Chiyoda Kenzai | Hitachi, Ltd. | Resid | 15,000 | October 1977 |
| Amagasaki | Nippon Oils & Fats | Hitachi, Ltd. | Resid | 20,000 | April 1978 |
| Sodegaura | Sumitomo Chemical | Mitsubishi H.I. | Resid | 300,000 | September 1976 |
| Sodegaura | Sumitomo Chemical | Sumitomo Chemical, Mitsubishi H.I. | Resid | 300,000 | October 1976 |
| Hirakatu | Kurabo | Kurabo | Resid | 30,000 | August 1975 |

TABLE 1.2.5-5. EXISTING FGT INSTALLATIONS OF SCR FIXED BED SYSTEMS OIL-FIRED INDUSTRIAL BOILERS[17]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Amagasaki | Kansai Paint | Hitachi, Ltd. | Distillate | 16,000 | October 1977 |
| Amagasaki | Nisshin Steel | Hitachi, Ltd. | Resid | 20,000 | August 1977 |
| Amagasaki | Nisshin Steel | Hitachi, Ltd. | Resid | 19,000 | July 1977 |
| Sakai | Nisshin Steel | Hitachi, Ltd. | Distillate | 30,000 | December 1978 |
| Hokkaichi | Shindaikyowa P.C. | Hitachi Zosen | Resid | 440,000 | November 1975 |
| Sodegaura | Sumitomo Chemical | Sumitomo Chemical Eng. | Resid | 30,000 | July 1973 |
| Sodegaura | Sumitomo Chemical | Sumitomo Chemical Eng. | Resid | 240,000 | March 1976 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 62,000 | October 1977 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 23,000 | December 1977 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 23,000 | June 1978 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 19,000 | July 1978 |
| Kawasaki | Nippon Yakin | Mitsubishi Kakoki | Resid | 14,000 | July 1978 |
| Kawasaki | Toho Gas | Mitsubishi Kakoki | Distillate | 30,000 | October 1977 |
| Chita | Toho Gas | Mitsubishi Kakoki | Distillate | 30,000 | October 1977 |

Table 1.2.5-6 shows the numerous industrial fixed packed bed SCR applications. Although gas-fired boilers, both industrial and utility, are numerous in Japan, few have been equipped with SCR units so far. This is due to the fact that less expensive $NO_x$ reduction by combustion modifications on these boilers has been adequate to meet environmental regulations. Fixed packed bed SCR systems are considered to be commercially available for natural gas-fired boilers at this time.

TABLE 1.2.5-6. GAS-FIRED SCR PLANTS IN JAPAN[17]

| Company | Site | Capacity ($Nm^3$/hr) | Reactor type | Completion date |
|---|---|---|---|---|
| Osaka Gas | Takaishi | 15,000x2 | FPB | December 1976 |
| Chubu Electric | Chita | 1,910,000 | FPB | April 1978 |
| Kyushu Electric | Kokura | 1,610,000 | FPB | July 1978 |
| Chubu Electric | Chita | 1,910,000 | FPB | September 1978 |
| Hyushu Electric | Kokura | 1,610,000 | FPB | December 1978 |

## 1.3 FLUE GAS TREATMENT FOR CONTROL OF $NO_x$ AND $SO_x$

Some FGT processes have the capability of removing $SO_x$ in addition to $NO_x$. These processes are typically more complex and costly than those which remove just $NO_x$; however, this is offset by the simultaneous dual pollutant control capability. For this reason, these processes are considered separately from the $NO_x$-only processes.

### 1.3.1 System Description

The following $NO_x$/$SO_x$ systems are considered for application to the coal-fired boilers:

- Parallel Flow SCR
- Adsorption
- Electron Beam Radiation
- Absorption-Reduction
- Oxidation-Absorption-Reduction
- Oxidation-Absorption.

Parallel flow SCR is selected as the only candidate for "best" $NO_x/SO_x$ system. The choice here is a combination of no serious secondary environmental impacts, system performance, system reliability, and status of development. The process is described briefly in Table 1.3.1-1.

TABLE 1.3.1-1. BEST $NO_x/SO_x$ EMISSION CONTROL SYSTEM FOR COAL-FIRED BOILERS

| Process | Description |
|---|---|
| Parallel Flow SCR | Utilizes $NH_3$ to catalytically reduce $NO_x$ after $SO_x$ is adsorbed by and reacted with catalyst; capable of achieving >90 percent $NO_x$ and $SO_x$ reduction; $SO_x$ saturated catalyst is regenerated while flue gas is diverted to alternate reactor. |

The $NO_x$ reduction reactions occurring in this process are the same as those described by reactions (1-1) and (1-2). The process utilizes an acceptor material to adsorb $SO_2$ and the product of the adsorption reaction then acts as an $NO_x$ reduction catalyst. Elemental copper is converted to oxide form by flue gas oxygen.[3]

$$Cu(s) + \tfrac{1}{2}O_2(g) \rightarrow CuO(s) \qquad (1\text{-}5)$$

Sulfur dioxide reacts with the copper oxide, as described by:[3]

$$SO_2(g) + \tfrac{1}{2}O_2(g) + CuO(s) \rightarrow CuSO_4(s) \qquad (1\text{-}6)$$

$SO_3$ in the flue gas is also removed:

$$SO_3(g) + CuO(s) \rightarrow CuSO_4(s) \qquad (1\text{-}7)$$

It is this final copper sulfate ($CuSO_4$) reaction product that acts as the primary catalyst for $NO_x$ reduction by ammonia in the parallel flow SCR $NO_x/SO_x$ system. After the spent reactor is isolated from flue gas flow, the reactor is purged with steam. A reducing gas, usually hydrogen, is then added which reacts with the copper sulfate in the following manner:[4]

$$CuSO_4(s) + 2H_2(g) \rightarrow Cu(s) + SO_2(g) + 2H_2O(g) \qquad (1\text{-}8)$$

The off-gas of this reaction is cooled to condense out the steam, reducing the gas volume and thus concentrating the $SO_2$. The concentrated $SO_2$ is compressed and sent to a workup section to produce either elemental sulfur, liquid $SO_2$, or sulfuric acid. (Sulfuric acid is produced in the cases studied in detail for this report.)

Besides the catalyst regeneration and sulfur treatment sections of the parallel flow SCR simultaneous $NO_x/SO_x$ system, the $NO_x$-only and $NO_x/SO_x$ parallel flow SCR systems are quite similar and the technical description contained in the previous section (1.2.1) applies here also.

No continuous coal-fired $NO_x$ removal test data for $NO_x/SO_x$ systems are available. Continuous oil-fired $NO_x$ removal test data for the parallel flow SCR $NO_x/SO_x$ system have been obtained from a 40 $MW_e$ unit in Japan. These data show the process capable of achieving up to 80 percent $NO_x$ control (average = 65 percent) and 90 percent $SO_x$ control. However, the process developers believe the process capable of achieving $NO_x$ control levels of greater than 90 percent as a result of testing on the 40 $MW_e$ unit. $SO_2$ removal results from coal-fired tests in the U.S. also show the process capable of attaining greater than 90 percent $SO_2$ reduction.

### 1.3.2 Economic Impacts

The annual and capital costs of this system applied to two coal-fired boilers and a residual oil-fired boiler are presented in this section. The costs are based on using a sulfuric acid plant for $SO_2$ workup and a compressor/gasholder for flow smoothing. Table 1.3.2-1 shows the annual costs of applications to coal-fired boilers. Two boiler types and two coals are presented and these data are plotted in Figure 1.3.2-1. Once again, economy of scale with large systems is evident; however, the effect is most significant for the high sulfur coal cases.

TABLE 1.3.2-1. ANNUAL COST OF PARALLEL FLOW $NO_x/SO_x$ CONTROL SYSTEMS

| Boiler | Fuel | Annual Cost, $1000/yr |
|---|---|---|
| Pulverized Coal | High Sulfur Eastern | 1805 |
| | Low Sulfur Western | 894 |
| Underfeed Stoker Coal | High Sulfur Eastern | 711 |
| | Low Sulfur Western | 462 |

Capital costs are shown in Figure 1.3.2-2 and the significant effect of boiler size on costs can again be seen.

Table 1.3.2-2 shows the annual cost of the single case studied. Only one case was analyzed for reasons described earlier and, as a result, it is not possible to plot the results or show trends.

TABLE 1.3.2-2. COSTS OF PARALLEL FLOW $NO_x/SO_x$ CONTROL SYSTEM

| Boiler | Fuel | Annual Cost, $1000/yr | Capital Cost, $1000 |
|---|---|---|---|
| Package, Watertube | Residual Oil | 1092 | 3801 |

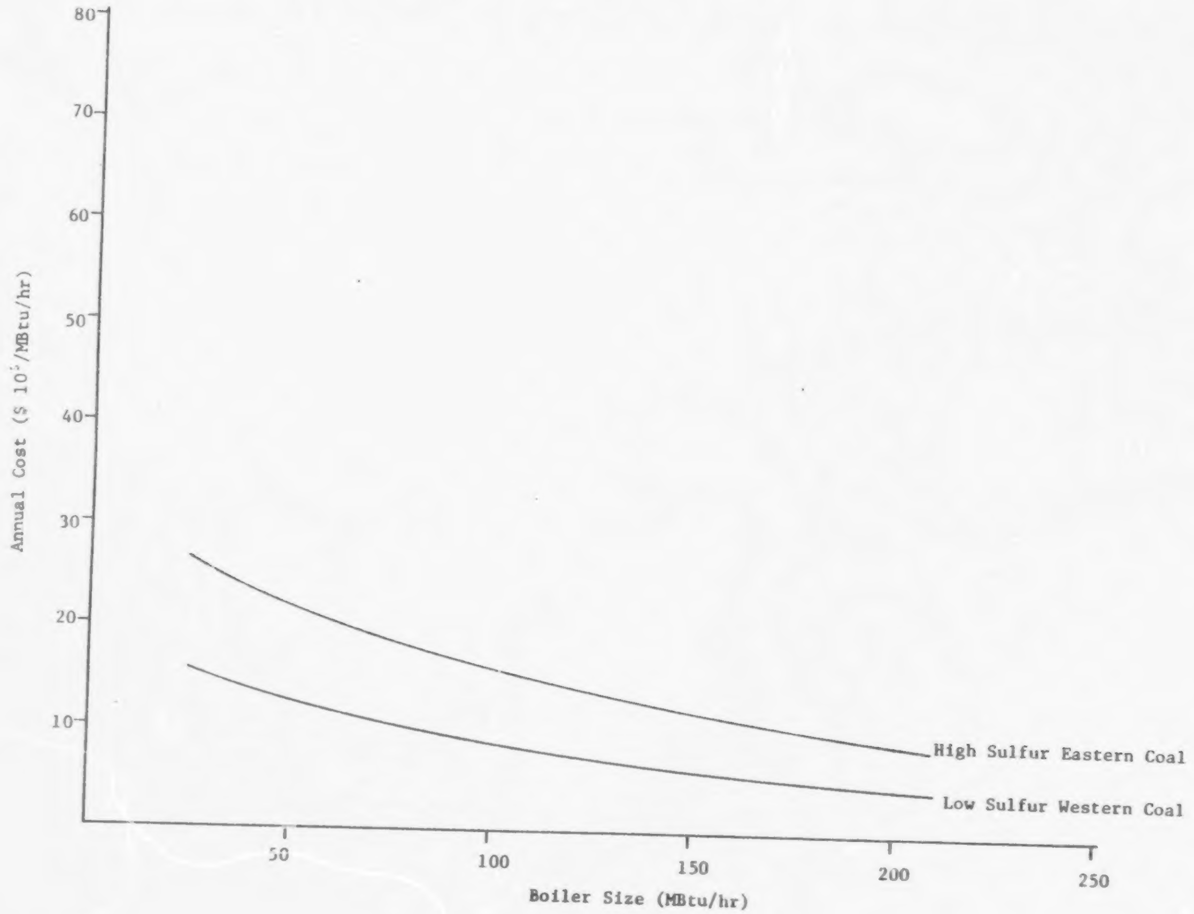

Figure 1.3.2-1. Annual cost of parallel flow SCR $NO_x/SO_x$ FGT for coal-fired boilers at intermediate level of control.

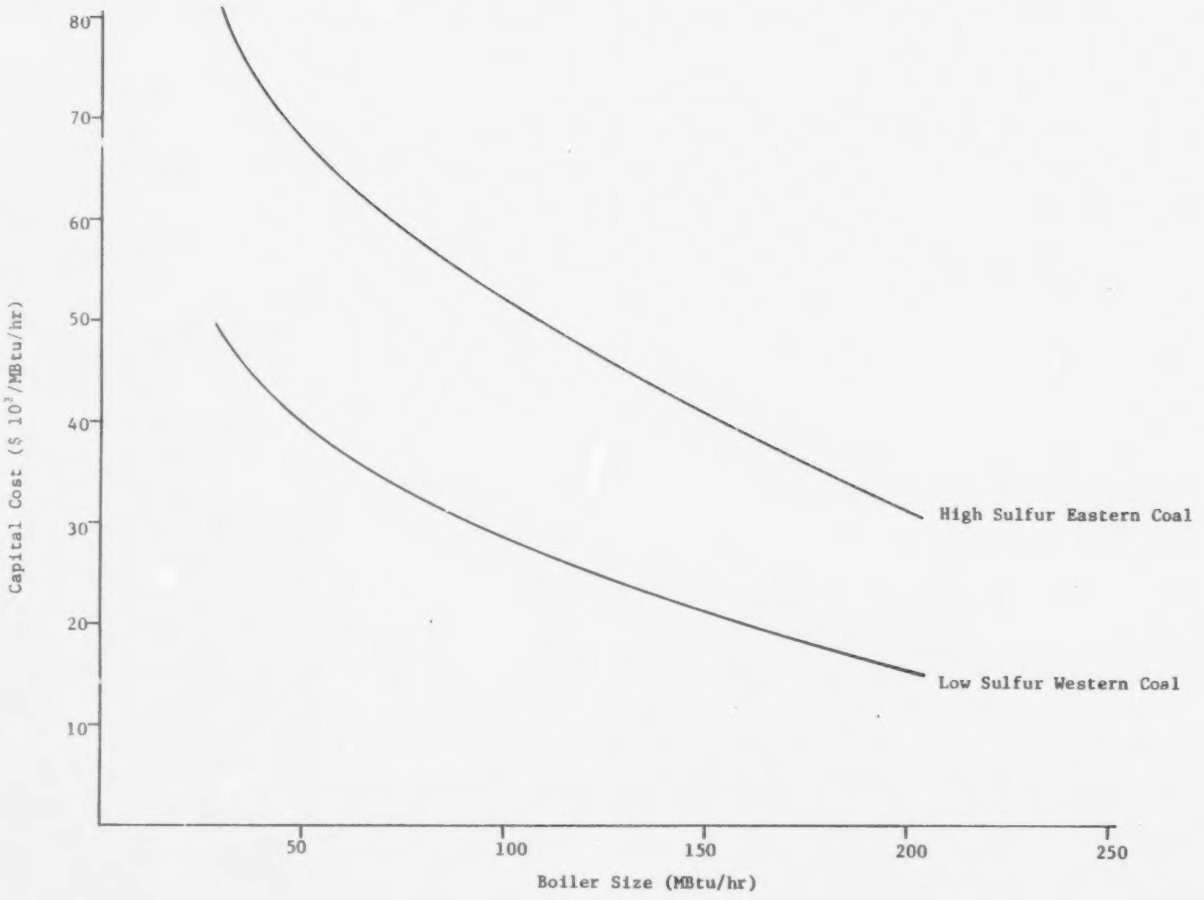

Figure 1.3.2-2. Capital cost of parallel flow SCR $NO_x/SO_x$ FGT for coal-fired boilers at intermediate level of control.

### 1.3.3 Energy Impacts

Energy usage for these cases is summarized in Table 1.3.3-1 and plotted in Figure 1.3.3-1 where energy consumption is plotted against flue gas flow rate. The curves are essentially linear with the high sulfur case having a significantly greater impact. With high sulfur coal, the energy usage is 7.7 percent of the boiler heat input for both boiler types. With the low sulfur coal, this figure drops to 2.2 percent of the boiler heat input.

TABLE 1.3.3-1. ENERGY CONSUMPTION OF $NO_x/SO_x$ CONTROL PROCESSES APPLIED TO COAL FIRED BOILERS

| Boiler | Fuel | Energy Consumption, $MW_t$ (MBtu/hr) |
|---|---|---|
| Pulverized Coal | High Sulfur Eastern | 4.5 (15) |
| | Low Sulfur Western | 1.2 (4.1) |
| Underfeed Stoker | High Sulfur Eastern | 0.68 (2.3) |
| | Low Sulfur Western | 0.20 (0.68) |

Energy use for the oil-fired boiler is sho[illegible]ble 1.3.3-2. Here again, it is not possible to plot the result.

TABLE 1.3.3-2. ENERGY CONSUMPTION OF PARALLEL FLOW $NO_x/SO_x$ CONTROL SYSTEM

| Boiler | Fuel | Energy Consumption | | |
|---|---|---|---|---|
| | | $MW_t$ | (MBtu/hr) | % of Boiler Heat Input |
| Package, Watertube | Residual Oil | 2.0 | (6.6) | 4.4 |

### 1.3.4 Environmental Impacts

The environmental impacts of this $NO_x/SO_x$ process are similar to those of the $NO_x$-only processes. The primary adverse environmental impact is from $NH_3$ emissions. The process developers claim that these emissions are low

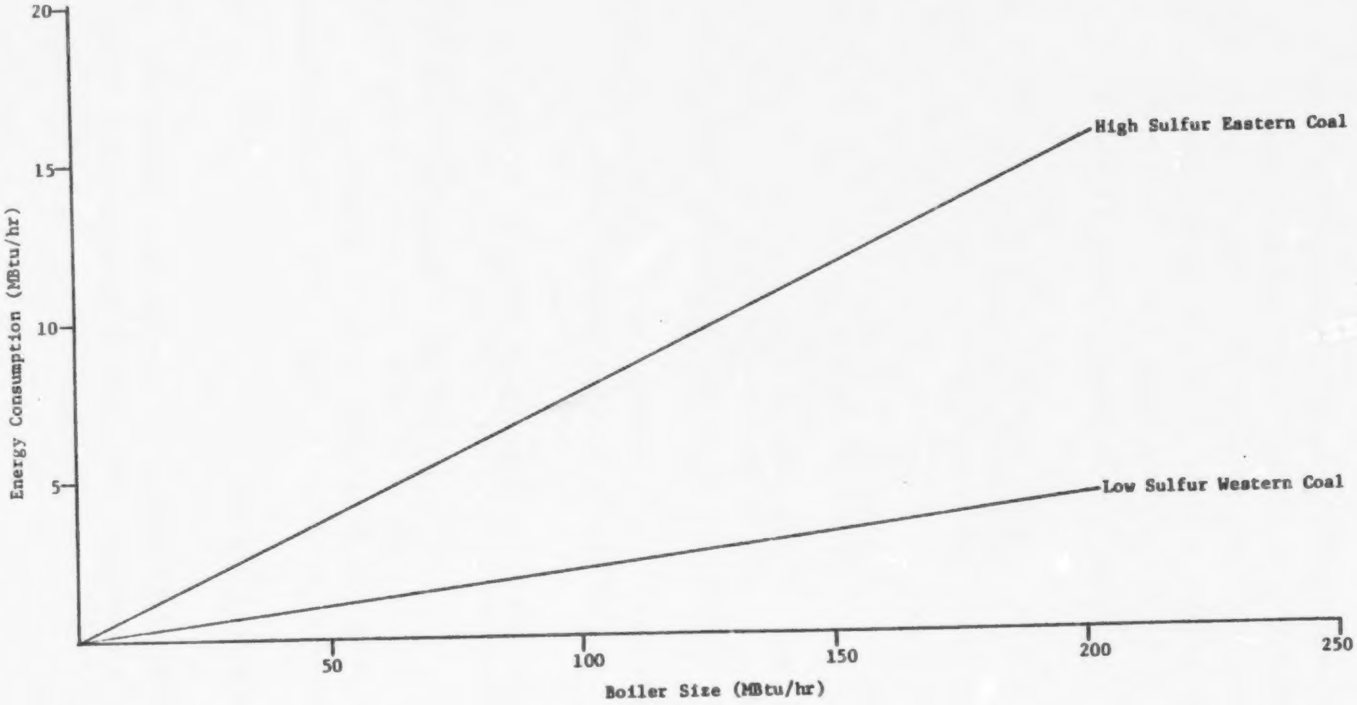

Figure 1.3.3-1. Energy consumption of parallel flow SCR $NO_x/SO_x$ FGT systems for coal-fired boilers.

(<10 ppm); however, a continuous $NH_3$ monitoring method for use with gases containing sulfur oxides will be necessary before $NH_3$ emissions can be accurately quantified. The potential problems with ammonium bisulfate and sulfate formation should be much less with the $NO_x/SO_x$ process since much of the $SO_x$ has been removed from the flue gas.

### 1.3.5 Development Status

The integrated process has been tested on oil but not coal; however, the $NO_x$ and $SO_x$ removal capabilities have been demonstrated separately. The $SO_2$ capabilities have been demonstrated on a pilot unit treating coal-fired flue gas. An EPA-sponsored U.S. test of the integrated process on flue gas from a coal-fired boiler is scheduled for 1980. Pilot and demonstration units of Shell/UOP process are summarized in Table 1.3.5-1 and commercial applications are summarized in Table 1.3.5-2.

TABLE 1.3.5-1. SHELL/UOP PROCESS, PILOT AND DEMONSTRATION UNIT

| Location/ Company | Designed By | Fuel/ Application | Size, $Nm^3/hr$ | Type of Operation | Dates |
|---|---|---|---|---|---|
| Shell Ref. at Pernis | Shell | Residual Fuel Oil-Proc. Heater | 600-1000 | $SO_x$-only | 1967-1972 |
| Rotterdam Utility | Shell | Coal-Steam Boiler | - | Heavy Fly Ash Loading | 1971 |
| Tampa Elec. Big Bend | UOP | Coal-Wet-Bottom Utility Boiler | 1200-2000 | $SO_x$-only<br>$SO_x$-$NO_x$ Simultaneous | 1974-1976 |
| JGC Yokohama Yokohama | JGC | Fuel Oil | 250-700 | $NO_x$-only | 1974- |
| Nippon Steel | JGC | Sintering Furnace | 2000 | $NO_x$-only | 1976-1978 |
| -- | JGC | Coke Oven | 400 | $NO_x$-only | 1976-1977 |

TABLE 1.3.5-2. SHELL/UOP PROCESS COMMERCIAL APPLICATIONS

| Unit | Designed By | Fuel/ Application | Size, $Nm^3/hr$ | Type of Operation | Dates |
|---|---|---|---|---|---|
| SYS* Yokkaichi | Shell | Residual Fuel Oil-Ref. Boiler | 125,000 | $SO_x$-only;<br>$NO_x$-$SO_x$ Simultaneous | 1973-1975<br>1975- |
| Kashima Oil Co. Ltd. | JGC | Fuel Oil-Process Unit Heater | 50,000 | $NO_x$-only | 1975- |
| Fuji Oil Co. Ltd. | JGC | CO Boiler | 70,000 | $NO_x$-only | 1976- |
| Nippon Steel Corp. | JGC | Sintering Furnace | 150,000 | $NO_x$-only | 1978- |

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3. Faucett, H.L., *et al*. Technical Assessment of $NO_x$ Removal Processes for Utility Application. EPA-600/7-77-127. November 1977. pp. 350.

4. *Ibid.*, p. 352.

5. *Ibid.*, p. 217.

6. Ando, Jumpei. $NO_x$ Abatement from Stationary Sources in Japan. EPA Report in Preparation. October 1978. pp. 3-67 - 3-82.

7. *Ibid.*, p. 3-7.

8. *Ibid.*, p. 4-41.

9. *Ibid.*, p. 4-95

10. Wong-Woo, Harmon. "Observation of FGD and Denitrification Systems in Japan." State of California Air Resources Board - SS-78-004. March 7, 1978. Appendix IV. p. 30.

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12. Faucett, H.L., *op cit.*, p. 5.

13. Ando, J., *op cit.*, p. 1-35.

14. $NO_x$ Control Review." Vol. 3, No. 4. Fall 1978. p. 3.

15. Ando, J., *op cit.*, pp. 4-1 - 4-133.

16. Noblett, J.G., *et al.* "Impact of $NO_x$ Selective Catalytic Reduction Processes on Flue Gas Desulfurization Processes," Draft Report. EPA Contract 68-02-2608, J.D. Mobley Project Officer, Radian Corporation. September 19, 1979.

17. Ando, J., *op cit.*, pp. 3-4, 3-5.

18. *Ibid.*, P. 3-7.

# SECTION 2

# EMISSION CONTROL TECHNIQUES

This section presents descriptions of all control techniques for $NO_x$ control by flue gas treatment (FGT). Each control technique is described separately, however, there may be several vendors offering processes that are similar. Where this occurs, an effort has been made to generalize the various processes into a single technique within a single category. This is usually, but not always, possible. Where significant differences exist, they are discussed separately.

A distinction has been made between those processes which remove only $NO_x$ and those which remove both $NO_x$ and $SO_2$. This is necessary because when final process comparisons are made it will be necessary to compare the cost of a $NO_x$ only process plus an FGD system versus the cost of a $NO_x/SO_x$ process. In the subsections which follow, all $NO_x$ only processes are grouped together and presented first and the $NO_x/SO_x$ process are presented second.

Economics for the various $NO_x$ control processes are presented only for comparison and use in Section III for process selection. These economic figures do not necessarily represent costs for application of these systems to industrial boilers, in fact, most were developed with utility applications in mind. However, at this time they are the only published economic data available. Detailed cost estimates for several systems as applied to industrial boilers will be developed for this study in Section IV.

## 2.1 Principles of Control

The FGT systems are examined on the basis of application to industrial boilers. These boilers are generally smaller than those used for utility applications and produce steam for purposes such as electrical power generation, process heating and space heating. They range in size from small package units to large field erected units. The demand on the boilers may be constant, such as with process heating, or cyclic, such as with space heating.[1] Industrial boilers generally have fewer burners than utility boilers and, therefore, taking just one burner out of service can have a significant effect on the flue gas characteristics.[2] Also, the stoker units typically run with higher excess air. These characteristics of industrial boilers indicate that typical flue gases can have a wide variety of characteristics.

This study considers seven standard boilers as selected for a variety of reasons in a separate study.[3] These boilers are described in Table 2.1-1. Three coals, two oils and natural gas are included as well as four sizes of coal-fired boilers. The coals considered are low sulfur western (0.6%S), low sulfur eastern (0.9%S) and high sulfur eastern (3.5%S). The two oils are distillate oil (#2) and residual oil (#6).

$NO_x$ is formed in boilers by two mechanisms. In one mechanism, thermal fixation, $N_2$ and $O_2$ present in the combustion air react to form NO. This reaction requires the high temperatures that are present in the burner flame and is dependent also on the $O_2$ concentration in the flame. The reaction does not reach equilibrium and therefore the amount of $NO_x$ formed by this mechanism is governed by reaction kinetics.[4] The second mechanism, fuel nitrogen conversion, involves the reaction of nitrogen contained in the molecular structure of the fuel with $O_2$ in the combustion air. The rate of reaction is a function of fuel nitrogen conversion and $O_2$ concentration. A more detailed description of the $NO_x$ formation mechanisms is contained in the Technology Assessment Report on $NO_x$ control by combustion modifications.

TABLE 2.1-1. CHARACTERISTICS OF THE STANDARD BOILERS CONSIDERED FOR ANALYSIS IN THIS REPORT

| Boiler | Fuel* | Size $MW_t$ | Control Level % |
|---|---|---|---|
| | $NO_x$-Only FGT Systems | | |
| Package, Firetube | Distillate Oil | 4.4 | 70, 90 |
| Package, Firetube | Natural Gas | 4.4 | 70, 90 |
| Package, Watertube | Residual Oil | 8.8 | 70, 90 |
| Package, Watertube Underfeed Stoker | LSW | 8.8 | 80 |
| Package, Watertube Chaingrate Stoker | LSW | 22 | 70, 80, 90 |
| Package, Watertube | Natural Gas | 44 | 70, 90 |
| Package, Watertube | Distillate Oil | 44 | 70, 90 |
| Package, Watertube | Residual Oil | 44 | 70, 90 |
| Field Erected, Watertube Spreader Stoker | LSW | 44 | 80 |
| Field Erected, Watertube Pulverized Coal | LSW | 58.6 | 70, 90 |
| | $NO_x/SO_x$ FGT Systems | | |
| Package, Watertube | Residual Oil | 44 | 80 $NO_x$, 85 $SO_x$ |
| Package, Watertube Underfeed Stoker | HSE<br>LSW | 8.8 | 80 $NO_x$, 85 $SO_x$ |
| Field Erected, Watertube Pulverized Coal | HSE<br>LSW | 58.6 | 80 $NO_x$, 85 $SO_x$ |

*HSE = High Sulfur Eastern Coal (3.5% S)
LSW = Low Sulfur Western Coal (0.6% S)

The $NO_x$ emissions for the various coals considered are different, presumably due to different fuel nitrogen concentrations. However, the emissions from the stoker boilers, on a ppm and mass per energy input basis, do not change from boiler to boiler. The mass rates do change due to differences in the flue gas flow rates for the various boilers. Emission rates for the standard boilers are shown in Table 2.1-2. The emission rates are based on AP-42 calculations.

In the sections which follow Section II, it is shown that $NO_x$ FGT system designs are not significantly affected by $NO_x$ concentration. The most significant design variables are flue gas flow rate and control level. For this reason, it is possible to generate information over the entire boiler size range while considering only one coal type. The coal chosen for analysis is low sulfur western since this coal has both the highest flue gas flow rates and $NO_x$ emissions.

FGT systems utilize either a gas phase reaction or liquid absorption to treat the flue gas. In most cases the gas phase reaction is between $NO_x$ and $NH_3$ in the presence of a solid phase catalyst. The catalyst is contained within a reactor and may be either fixed or moving bed. The $NO_x$ is converted to $N_2$ which exits with the flue gas.

Systems utilizing a liquid absorption technique contact flue gas and absorbent in conventional scrubbers. The absorbed $NO_x$ either remains in the scrubbing liquor and is treated in the liquid phase or reacts with a solute to form $N_2$ which degasses and leaves with the flue gas.

The $NO_x$ FGT systems discussed in the following subsections are divided into two categories. Those which remove only $NO_x$ are presented first and the simultaneous $NO_x/SO_x$ processes are discussed second. The distinction is made since the two process types cannot be accurately compared unless FGD flue gas desulfurization (FGD) is included with the $NO_x$-only processes. This comparison will be made, but only in the Comprehensive Technology Assessment

TABLE 2.1-2. $NO_x$ EMISSION RATES FOR THE STANDARD BOILERS

| Boiler | Fuel* | $NO_x$ Emissions | | | | |
|---|---|---|---|---|---|---|
| | | g/s | (lb/hr) | g/MJ | (lb/$10^6$Btu) | ppm |
| Package, Firetube | Distillate Oil | 0.300 | (2.38) | 0.0688 | (0.16) | 97 |
| Package, Firetube | Natural Gas | 0.332 | (2.63) | 0.0774 | (0.18) | 104 |
| Package, Watertube | Residual Oil | 2.02 | (16.0) | 0.228 | (0.53) | 373 |
| Package, Watertube Underfeed Stoker | HSE | 2.40 | (19.05) | 0.275 | (0.64) | 335 |
| | LSE | 2.06 | (16.35) | 0.237 | (0.55) | 288 |
| | LSW | 2.95 | (23.40) | 0.335 | (0.78) | 402 |
| Package, Watertube Chaingrate | HSE | 6.02 | (47.70) | 0.275 | (0.64) | 336 |
| | LSE | 5.15 | (40.80) | 0.232 | (0.54) | 290 |
| | LSW | 7.40 | (58.65) | 0.335 | (0.78) | 401 |
| Package, Watertube | Natural Gas | 3.31 | (26.26) | 0.0753 | (0.175) | 110 |
| Package, Watertube | Distillate Oil | 2.99 | (23.76) | 0.0680 | (0.158) | 101 |
| Package, Watertube | Residual Oil | 7.47 | (60.00) | 0.172 | (0.40) | 292 |
| Field Erected, Watertube Spreader Stoker | HSE | 12.0 | (95.40) | 0.275 | (0.64) | 337 |
| | LSE | 10.3 | (81.45) | 0.232 | (0.54) | 288 |
| | LSW | 14.8 | (117.15) | 0.335 | (0.78) | 400 |
| Field Erected, Watertube Pulverized Coal | HSE | 19.2 | (152.46) | 0.327 | (0.76) | 466 |
| | LSE | 16.5 | (130.50) | 0.280 | (0.65) | 396 |
| | LSW | 23.7 | (187.56) | 0.404 | (0.94) | 550 |

*Coal types: HSE = High Sulfur Eastern (3.5%S)
LSE = Low Sulfur Eastern (0.9%S)
LSW = Low Sulfur Western (0.6%S)

Report (CTAR) which follows completion of the Individual Technology Assessment Reports (ITAR's). Therefore, in Section III of this ITAR, $NO_x$-only processes will only be compared with other $NO_x$-only processes and $NO_x/SO_x$ processes will only be compared with other $NO_x/SO_x$ processes. This distinction will be maintained throughout the other sections of the ITAR also. The $NO_x$-only processes described are:

- Fixed Packed Bed Selective Catalytic Reduction (SCR)
- Moving Bed SCR
- Parallel Flow SCR
- Absorption-Oxidation

The $NO_x/SO_x$ processes described are:

- Parallel Flow SCR
- Adsorption
- Electron Beam Radiation
- Absorption-Reduction
- Oxidation-Absorption-Reduction
- Oxidation-Absorption

## 2.2 CONTROLS FOR COAL-FIRED BOILERS

### 2.2.1 Selective Catalytic Reduction-Fixed Packed Bed Reactors

Fixed packed bed systems for selective catalytic reduction of $NO_x$ are applicable only to flue gas streams containing particulate emissions of less than 20 $mg/Nm^3$. Particulate emissions for all coals are higher, on the order of 1-5 grams per $Nm^3$. Although it is possible to install a hot ESP to reduce the particulate level to 20 $mg/Nm^3$ this is expensive and not always effective. For these reasons fixed packed bed SCR systems are not considered for application to coal-fired boilers by process vendors.[5]

### 2.2.2 Selective Catalytic Reduction-Moving Bed Reactors

Moving bed systems for selective catalytic reduction of $NO_x$ are applicable only to flue gas streams containing less than 1 g/Nm$^3$. Particulate emissions for all coals are higher, on the order of 1-5 grams per Nm$^3$. Although it is possible to install a hot ESP to reduce the particulate level to 1 g/Nm$^3$ this is expensive and not always effective. For these reasons moving bed SCR systems are not considered for application to coal-fired boilers in this report.

### 2.2.3 Selective Catalytic Reduction-Parallel Flow Reactor

#### 2.2.3.1 System Description--

The distinguishing aspect of this process is the catalyst shape which is produced in a variety of shapes. The catalysts are produced in either a honeycomb, pipe, or plate shape. Both metal and ceramic supports are employed. Several shapes are illustrated in Figure 2.2.3-1. The catalyst shapes allow particulate laden flue gas to pass through the reactor with no inertial impaction of the particles while the $NO_x$ is transported to the catalyst surfaces by basic diffusion. The catalysts can handle all of the particulate levels emitted by the standard boilers. All of the catalysts considered here for use in treating flue gas containing $SO_2$ and $SO_3$ are resistant to poisoning by these compounds. Long term tests of these catalysts in the presence of $SO_x$ have shown very little or no decrease in activity or selectivity.

The reactors used are similar to standard fixed packed bed units and an example is shown in Figure 2.2.3-2. The catalyst is usually prepared in small modules and manually stacked within the reactor. The specific arrangement will depend on the particular process under consideration.

A typical flow diagram for a parallel flow SCR system is shown in Figure 2.2.3-3. The arrangement is similar to the other SCR processes in that hot flue gas leaving the boiler economizer is injected with $NH_3$ and

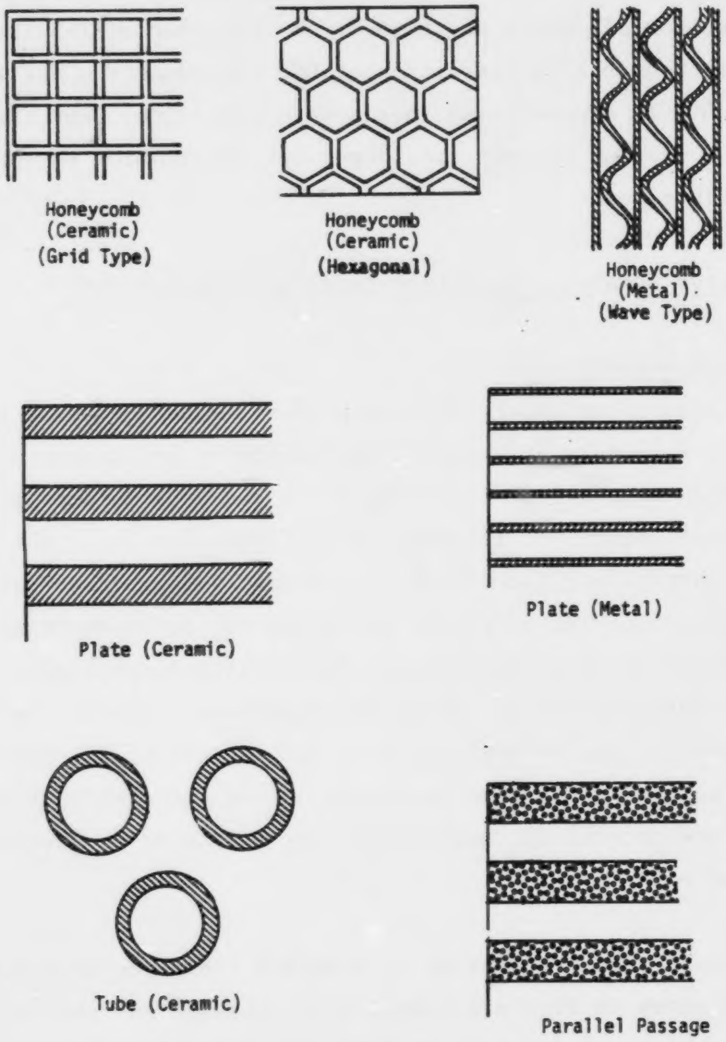

Honeycomb (Ceramic) (Grid Type)
Honeycomb (Ceramic) (Hexagonal)
Honeycomb (Metal) (Wave Type)
Plate (Ceramic)
Plate (Metal)
Tube (Ceramic)
Parallel Passage

Figure 2.2.3-1. Shapes of parallel flow catalysts.[22]

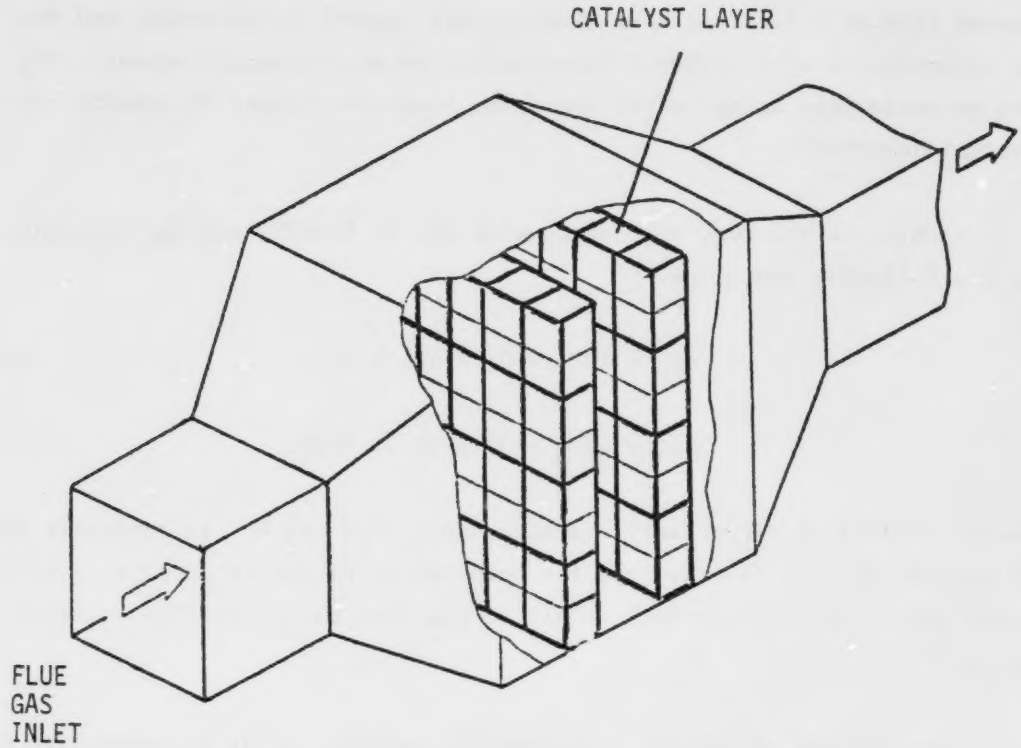

CATALYST LAYER
FLUE
GAS
INLET

Figure 2.2.3-2. Typical reactor used with parallel flow SCR process.[23]

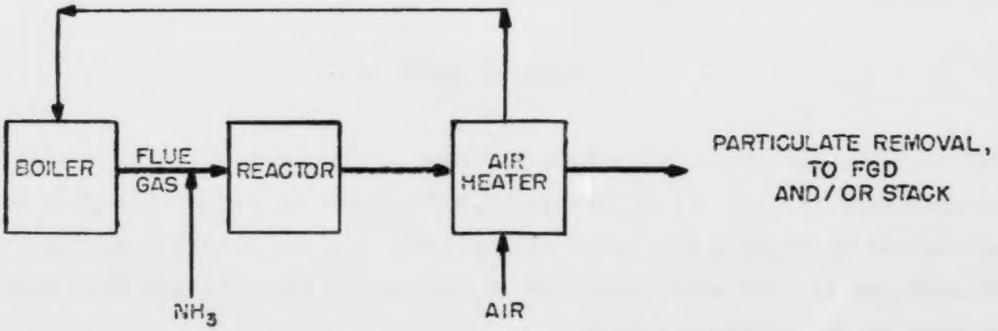

BOILER
FLUE
GAS
REACTOR
AIR
HEATER
PARTICULATE REMOVAL,
TO FGD
AND / OR STACK
$NH_3$
AIR

Figure 2.2.3-3. Flow diagram for parallel flow SCR process.[24]

passed through a catalyst bed. Temperature control is important and can be accomplished with either a fired heater or an economizer bypass. $NH_3$ can be controlled using boiler operating condition inputs to conventional control components.

Within the reactor, $NO_x$ reacts with $NH_3$ to form $N_2$ and $H_2O$ according to the following reactions.[12]

$$4NH + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \qquad (2\text{-}1)$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \qquad (2\text{-}2)$$

Reaction (2-1) is the primary reaction since flue gas $NO_x$ is typically 90-95 percent NO. $O_2$ is necessary for both reactions and is present in sufficient quantities (>3 percent) in all of the flue gases from the standard boilers.

The catalyst volume for a desired $NO_x$ removal can be determined by the fundamental design equation for a plug flow reactor.[13]

$$\frac{V}{F} = \int_o^x \frac{dx}{r} \qquad (2\text{-}3)$$

The reaction rate, r, can be expressed as

$$r = k[NH_3]^a\ [NO]^b\ [O_2]^c \qquad (2\text{-}4)$$

The variables presented here have the same definitions as those presented in equations 2-3 and 2-4 of Section 2.3.2. Catalyst volume can also be determined by knowing the space velocity for a given catalyst and NO conversion level. The space velocity is defined as the flue gas flow rate divided by the catalyst volume.

The reaction rate is different for each catalyst formulation since different catalysts will lower the activation energy by different amounts. The activation energy affects the reaction rate constant, k, according to the Arrhenius equation.

$$k = Ae^{-\frac{E}{RT}} \tag{2-5}$$

Example values of k, a, b, and c for two catalyst formulations are shown in Table 2.2.3-1.

An important design variable with catalytic systems is the space velocity which expresses the volume of catalyst required to treat one volume per hour of flue gas. Space velocity varies with catalyst formulation, catalyst shape, and control level. Typical values of space velocity for various catalyst shapes are shown in Table 2.2.3-2. Also shown are other catalyst design variables such as catalyst dimensions, gas velocities, bed depth and pressure drop. Ranges of values are used since specific values are different for each catalyst. The values shown pertain to 90 percent $NO_x$ removal and an $NH_3/NO_x$ mole ratio of 1:1.

Both $NH_3/NO_x$ ratio and space velocity will change with removal level. The $NH_3/NO_x$ mole ratio will range from 0.7-1.0 and the space velocity will range approximately as shown in the table for control levels of 70 to 90 percent.[15]

Variables associated with the boiler can also affect the performance of these systems. These are

- flue gas flow rate
- $NO_x$ concentration
- boiler load variability

TABLE 2.2.3-1. REACTION RATE DATA FOR TWO CATALYST FORMULATIONS[11]

Catalyst: $V_2O_5$ on $Al_2O_3$

$$k = 2.05 \times 10^3 e^{-\frac{9650}{RT}}$$

$a = 0.30$

$b = 0.22$

$c = 0.05$

Catalyst: Fe-Cr on $Al_2O_3$

$$k = 3.25 \times 10^3 e^{-\frac{10,860}{RT}}$$

$a = 0.45$

$b = 0.10$

$c = 0.15$

TABLE 2.2.3-2. CATALYST DESIGN VARIABLES FOR VARIOUS CATALYST SHAPES[25] (Basis: 90% $NO_x$ removal at $NH_3/NO_x$ ratio of 1:1, 350-400°C)

| | Honeycomb (metallic) | Honeycomb, tube (ceramic) | Parallel Plate | |
|---|---|---|---|---|
| | | | (Ceramic) | (Metallic) |
| Catalyst size (mm) | | | | |
| Thickness | 1 | 2.3-5 | 8-10 | 1 |
| Opening | 4-8 | 6-20 | 8-14 | 5-10 |
| Gas velocity (m/sec)[a] | 2-6 | 5-10 | 5-10 | 4-8 |
| Bed depth (m) | 1-2 | 1.5-5 | 4-6 | 2-5 |
| SV (1,000 $hr^{-1}$)[b] | 5-8 | 4-8 | 1.5-3 | 2-4 |
| Pressure drop ($mmH_2O$) | 40-80 | 40-160 | 80-160 | 60-120 |

[a] Velocity at 350-400°C in open column (superficial velocity).

[b] Gas volume ($Nm^3$/hr)/catalyst bed volume ($m^3$).

The flue gas flow rate and control level determine the catalyst volume (hence reactor size) necessary. Increases in either also increase the reactor size. The $NO_x$ concentration is primarily a function of fuel type used in the standard boilers. Higher concentrations require larger $NH_3$ storage and vaporization equipment; reactor size is not affected. Boiler load can affect several things including flue gas temperature, flow rate, and $NO_x$ concentration. It is necessary to maintain reaction temperatures of 350 to 400°C. Temperature control equipment may be necessary to accomodate large boiler load variations which may lower the flue gas temperature. Where these variations are present, some equipment overdesign may be warranted to insure a constant control level. These variables are discussed in more detail in the section on moving bed SCR systems for coal-fired boilers, Section 2.2.2.

Parallel flow SCR processes have been applied in Japan to several residual oil-fired industrial boilers. Oil-fired utility boilers and other sources with high particulate concentrations are also being treated. Two applications to coal-fired utility boilers are planned for 1980 (Table 2.2.3-3) although none exist at the present time. A coal-fired pilot unit demonstration of one $NO_x$-only parallel flow design is currently underway in the U.S. under EPA sponsorship and several have been conducted in Japan. The EPA facility should be operational by mid-1979. Also, a parallel flow pilot system will be applied to flue gas from a coal-fired boiler in a study sponsored by the Electric Power Research Institute (EPRI). The unit is expected to be operational by 1980. A list of vendors of parallel flow SCR systems is presented in Table 2.2.3-4. The number of pilot unit demonstrations indicates that application of parallel flow SCR processes to coal-fired industrial boilers is feasible.

### 2.2.3.2 System Performance--

Performance data based on pilot plant testing were not found in the literature, however, data do exist for oil-fired applications. Since many of the flue gas characteristics are similar for oil and coal-fired boilers,

TABLE 2.2.3-3. PLANNED FGT INSTALLATIONS OF SCR COAL-FIRED UTILITY BOILERS[26]

| Location | User | Process Developer | Fuel | Capacity ($Nm^3$/hr) | Completion Date |
|---|---|---|---|---|---|
| Takehara | Electric Power C. | Not yet announced | Coal | 800,000 | July 1981 |
| Tomato | Hokkaido Electric | Hitachi, Ltd. | Coal | 88,000 | October 1980 |

TABLE 2.2.3-4. PROCESS VENDORS OF PARALLEL FLOW SCR PROCESSES[28]

| Vendor | Demonstrated on Coal Yes/No | Demonstrated on Coal Scale |
|---|---|---|
| Hitachi Zosen | yes | pilot |
| Hitachi, Ltd. | yes | pilot |
| Japan Gasoline Corp. | no | -- |
| Mitsui Engineering & Shipbuilding | no | — |
| Mitsubishi Heavy Industries | yes | pilot |
| Ishikawajima-Harima Heavy Industries | yes | pilot |
| Kobe Steel | no | -- |
| Kawasaki Heavy Industries | yes | pilot |
| Shell/UOP | by 1979 | pilot |

it is expected that the FGT performance will be roughly similar. Detailed data on oil-fired applications are contained in Section 2.3.

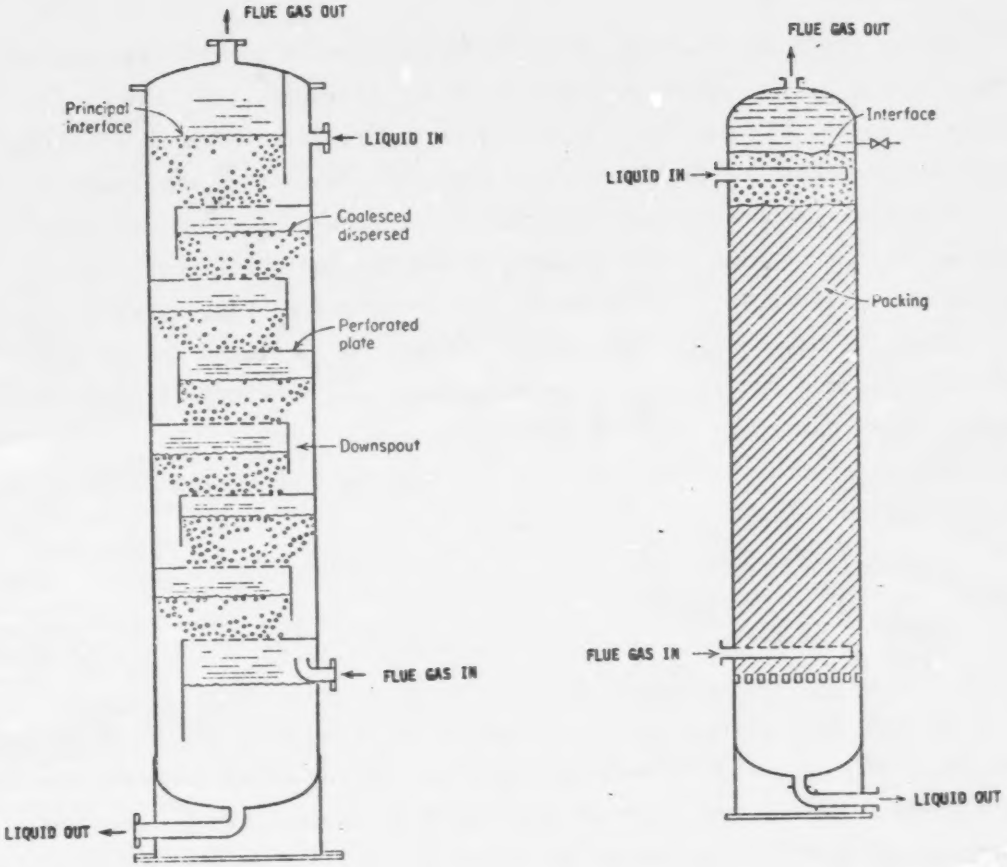
There are some potential problems downstream of the SCR systems due to the presence of the unreacted ammonia in the flue gas. Two things can happen: 1) the $NH_3$ can react with $SO_2$ or $SO_3$ to form ammonium bisulfate or ammonium bisulfate or 2) the $NH_3$ can enter the downstream equipment unreacted. The bisulfate has been shown to cause air preheater pluggage and this is the subject of ongoing research both at the EPA and the Electric Power Research Institute (EPRI). Both the bisulfate and sulfate exist as a particulate, but may be difficult to collect if the particles are submicron in size. Unreacted $NH_3$ is not likely to present any operational problems. A recent study has shown that if an ESP exists downstream, then most of the $NH_3$ will exit with the ash. $NH_3$ can actually improve the performance of an FGD system.[129]

## 2.2.4 Absorption-Oxidation

### 2.2.4.1 System Description--

Absorption-oxidation processes remove $NO_x$ from flue gas by absorbing the NO or $NO_x$ into a solution containing an oxidant which converts the $NO_x$ to a nitrate salt. Two types of gas/liquid contactors can be used and examples of each type are shown in Figure 2.2.4-1. Both perforated plate and packed towers accomplish $NO_x$ absorption by generating high gas/liquid interfacial areas. The choice of one type of contactor is a design decision made to achieve a given removal for the least cost.

A generalized process flow diagram is shown in Figure 2.2.4-2. Flue gas is taken from the boiler after the air preheater. Before the gas can be sent to the $NO_x$ absorber, it must be $SO_2$-free since $SO_2$ consumes prohibitive amounts of the costly liquid-phase oxidant. In most cases, the oxidant is permanganate ($MnO_4^-$), but $Ca(ClO)_2$ can also be used. Therefore, a conventional FGD unit is required ahead of the $NO_x$ absorber. A prescrubber to cool


FLUE GAS OUT
Principal interface
LIQUID IN
Coalesced dispersed
Perforated plate
Downspout
FLUE GAS IN
LIQUID OUT
FLUE GAS OUT
Interface
LIQUID IN
Packing
FLUE GAS IN
LIQUID OUT

Perforated Plate Absorber

Packed Absorber

Figure 2.2.4-1. Gas/liquid contactor options for Absorption-Oxidation Processes.[29]

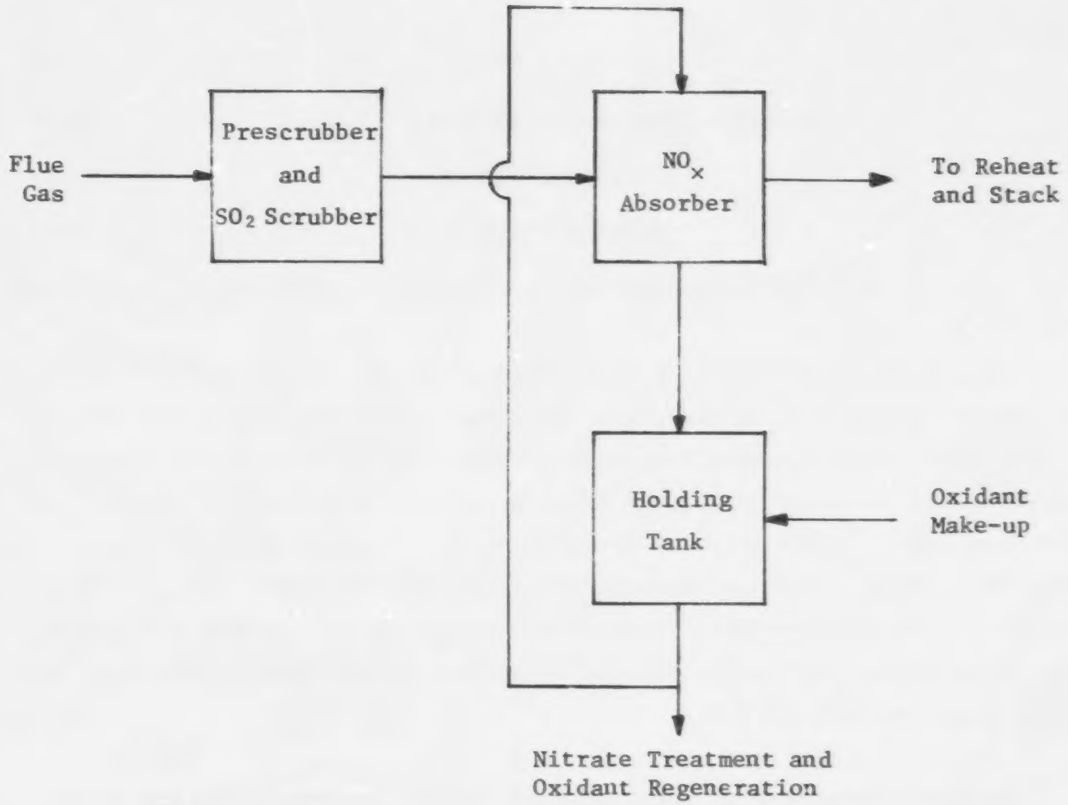

Flue Gas
Prescrubber and $SO_2$ Scrubber
$NO_x$ Absorber
To Reheat and Stack
Holding Tank
Oxidant Make-up
Nitrate Treatment and Oxidant Regeneration

Figure 2.2.4-2. Process flow diagram for Absorption-Oxidation Process.[30]

the gas and remove both particulates and $Cl^-$ prior to FGD is also necessary. After having passed through these two scrubbing sections, the flue gas enters the distributing space at the bottom of the $NO_x$ absorber, below the packing or plates. The gas passes upward through the column, countercurrent to the flow of the liquid absorbent/oxidant (usually a KOH solution containing $KMnO_4$). $NO_x$ is absorbed and then oxidized over the length of the column according to the following reactions.[31]

$$NO\,(g) \rightarrow NO(aq) \tag{2-6}$$

$$NO(aq) + KMnO_4(aq) \rightarrow KNO_3(aq) + MnO_2(s) \tag{2-7}$$

$$2NO_2(g) \rightarrow N_2O_4(g) \tag{2-8}$$

$$N_2O_4(g) \rightarrow N_2O_4(aq) \tag{2-9}$$

$$N_2O_4(aq) + 2K_2MnO_4(aq) \rightarrow 2KMnO_4(aq) + 2KNO_2(aq) \tag{2-10}$$

Since most of the $NO_x$ from combustion processes occurs as NO,[32] reactions 2-6 and 2-7 predominate. The clean gas passes out of the top of the absorber to a heater for plume buoyancy and is sent to the stack. The absorbing solution drops to a holding tank where makeup KOH and/or $KMnO_4$ are added. This solution flows to a centrifuge to separate the solid $MnO_2$ which is then electrolytically oxidized to $MnO_4$. The remaining solution is either concentrated in an evaporator to form a weak $KNO_3$ solution or is electrochemically treated to produce a weak $HNO_3$ solution and a mixed stream of KOH and $KNO_3$.

The fundamental design equation used for gas absorption column design is

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \tag{2-11}$$

where $y$ = bulk $NO_x$ concentration (mole fraction) of gas phase at any given point in column

$y-y^*$ = overall driving force for absorption ($y^*$ being the $NO_x$ concentration of a gas in equilibrium with a given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/(ft$^2$)(hr)(mole fraction)

$a$ = area of gas-liquid interface per unit packed volume, ft$^2$/ft$^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/(ft$^2$)(hr)

$Z$ = length of packed section of column, ft

In a column containing a given packing or plate configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity, obviously, must be lower than the flooding velocity. How much lower is a design decision. Often, it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[34] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter is a major problem for these systems trying to achieve large $NO_x$ reductions since NO is relatively insoluble in water. This can be seen in Table 2.2.4-1.

TABLE 2.2.4-1. NITROGEN OXIDES CHARACTERISTICS[35]

| | Boiling Point, °C | Solubility in Cold Water (0°C), cm$^3$ | Solubility in Hot Water (60°C), cm$^3$ |
|---|---|---|---|
| NO | -151.8 | 7.34/100 cc $H_2O$ | 2.37/100 cc $H_2O$ |
| $NO_2$ | 21.2 | soluble, decomposes | -- |

One can see that NO has a very limited solubility in water and, since most $NO_x$ is present as NO, the rate of mass transfer (absorption) is going to be relatively slow. This means that the absorber must be tall with a high liquid flow rate. Table 2.2.4-2 presents the effects of boiler/flue gas variables on the design of absorption-oxidation systems.

TABLE 2.2.4-2. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Requires FGD pretreatment |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased oxidant concentration |

Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load changes. The column size and the liquid and oxidant flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

None of the sources consulted for this study could supply typical ranges for operating variables such as liquid/gas ratio, reagent concentrations or pressure drops and, as a result, none are presented here. Economic data were not presented either. One source did estimate the removal for absorption-oxidation processes to be 85 percent.[36]

Presently, absorption-oxidation processes are still in the pilot unit stage of development. Table 2.2.4-3 presents a list of absorption-oxidation process vendors and the status of development of their projects. One can see from the table that no coal-fired flue gas tests have been performed.

TABLE 2.2.4-3. PROCESS VENDORS OF ABSORPTION-OXIDATION PROCESSES[37,38]

| Vendor | Status of Development |
|---|---|
| Hodogaya | No information available; stopped development on process |
| Kobe Steel | 1974: 1000 $Nm^3$/hr gas from iron-ore sintering furnace; stopped development on process |
| MON (Mitsubishi Metal, MKK, Nikon Chemical) | 1974: 4000 $Nm^3$/hr flue gas from oil-fired boiler |
| Nissan Engineering | 1972: 4 pilot plants, 100-2000 $Nm^3$/hr tail gas from $HNO_3$ plant |

### 2.2.4.2 System Performance--

No coal-fired tests have been made. No information has been published on tests conducted with other fuels. The relative insolubility of NO in water may present a major obstacle to achieving the stringent level of control (90 percent $NO_x$ reduction) by absorption-oxidation processes. Another primary drawback of absorption-oxidation systems is the production of nitrate salts (see Equation 4-2), a secondary pollutant. These processes probably could not be applied on a large scale as wastewater treatment systems (chemical or biological) do not remove nitrogen compounds from the wastewater.[39] Trying to recover the nitrates as nitric acid for industrial use or potassium nitrate for fertilizer does not seem promising as the by-products are of low quality. Also, the use of an expensive, liquid-phase oxidant requires stainless steel and other corrosion resistant materials of construction. High sulfur coals require an FGD system prior to the $NO_x$ absorber to prevent excessive oxidant consumption by $SO_2$. The process steps of several absorber columns in series (large fan requirements), oxidant regeneration (electrolysis), and flue gas reheat (inline heater) are all energy intensive and present technical and economic disadvantages.

## 2.2.5 Selective Catalytic Reduction-$NO_x/SO_2$ Removal

### 2.2.5.1 System Description--

From a $NO_x$ removal standpoint, this process is very similar to those discussed in Sections 2.2.2 and 2.2.3. The primary difference is the additional equipment necessary to collect and process the $SO_2$. The main feature of the process is the reactor and catalyst which remove both $NO_x$ and $SO_2$. This process was developed by Shell although the U.S. licensor, UOP, is currently marketing and developing the process. The $NO_x/SO_2$ version of the process is commonly called the SFGT process which stands for the Shell Flue Gas Treatment process.

A uniquely designed parallel flow type of reactor is used to avoid problems with particulates. The reactor consists of a series of packages containing catalyst material, arranged in a parallel configuration which allows flue gas flow between the packages. Each package consists of catalyst material placed between two layers of wire gauze. Figure 2.2.5-1 illustrates the internals of the parallel passage reactor. The flue gas flows between the catalyst packages and not directly through the catalyst material. This prevents plugging of the catalyst with particulate matter in the flue gas.

For convenient fabrication and handling, catalyst packages of a standard size are appropriately spaced and placed in a container to form a unit cell or module. $SO_2$ removal efficiency and capacity are determined by the number of unit cells placed in series in a cell stack. For a given level of $SO_2$ removal, a greater number of cells in the stack increases the capacity and reduces the frequency of regeneration. The number of stacks is determined largely by the flue gas rate and the flue gas velocity through a single stack is generally not a design variable. For most design situations, 4 to 5 unit cells in a stack are adequate to achieve high $SO_2$ removal, however, if a high level of denitrification is required, more unit cells per stack may be necessary.

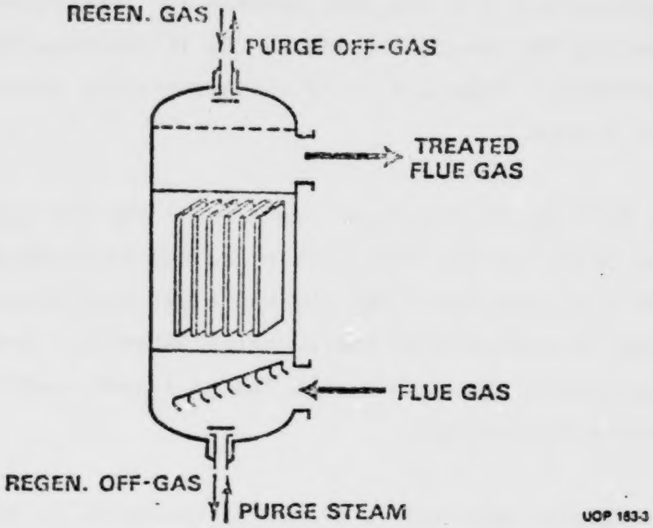


Figure 2.2.5-1. The SFGT parallel flow reactor.[40]

The SFGT process is a dry process with two or more reactors operating in a cyclic manner. The desulfurization aspect of the process is regenerable, while $NO_x$ removal is accomplished by catalytic reduction with ammonia. The catalyst material is commonly called an acceptor since $SO_2$ removal involves adsorption or "acceptance" of $SO_2$. The desulfurization cycle consists of the following steps:

1) oxidation of acceptor bed/acceptance of $SO_2$,
2) purge reactor,
3) regeneration with reducing gas, and
4) purge reactor.

The products of the oxidation and acceptance reactions in step 1 above catalyze the reaction of $NO_x$ with ammonia to form nitrogen and water. $NO_x$ removal is accomplished by metering ammonia into the untreated flue gas upstream of the reactors. The catalytic reaction takes place across the partially spent acceptor beds.

Also associated with the SFGT process are facilities for generating reducing gas and for the processing of $SO_2$ in regeneration off gases into sulfur by-products. Figure 2.2.5-2 illustrates the process flow for a typical SFGT system.

Boiler flue gas is withdrawn upstream of the air preheater and particulate removal device by the SFGT system fan and discharged to the reactor inlets. The flue gas then flows through fixed bed reactors in open channels alongside and in contact with the acceptor material. Ammonia is added to the flue gas upstream of the SFGT system fan to insure complete mixing before the flue gas enters the reactor.

Fresh acceptor material is elemental copper on an alumina support. This is converted to the oxide form by flue gas oxygen shortly after initiation of the acceptance cycle. $SO_2$ is removed by reaction with the copper oxide and oxygen as the flue gas flows through the channels, converting the acceptor material to copper sulfate. Simultaneous with the desulfurization process, the reduction of flue gas $NO_x$ by ammonia is selectively catalyzed by copper oxide and copper sulfate in the acceptor bed. As the flue gas leaves the SFGT system reactors it is returned to the boiler flue gas duct downstream of SFGT fan suction.

Flue gas is fed to a reactor until an unacceptable amount of $SO_2$ begins to pass through the reactor. This occurs when a large fraction of the acceptor has been converted to the sulfate form. Flue gas flow is then diverted to another reactor and the spent reactor is isolated. Any flue gas remaining in the spent reactor is purged with an inert gas such as steam, and the regeneration cycle is initiated.

Regeneration is accomplished by passing a reducing gas through the bed countercurrent to the direction of the flue gas flow. The reducing gas, which is primarily hydrogen, reacts with the copper sulfate in the spent reactor to convert it to elemental copper. An off gas of $SO_2$ and water is

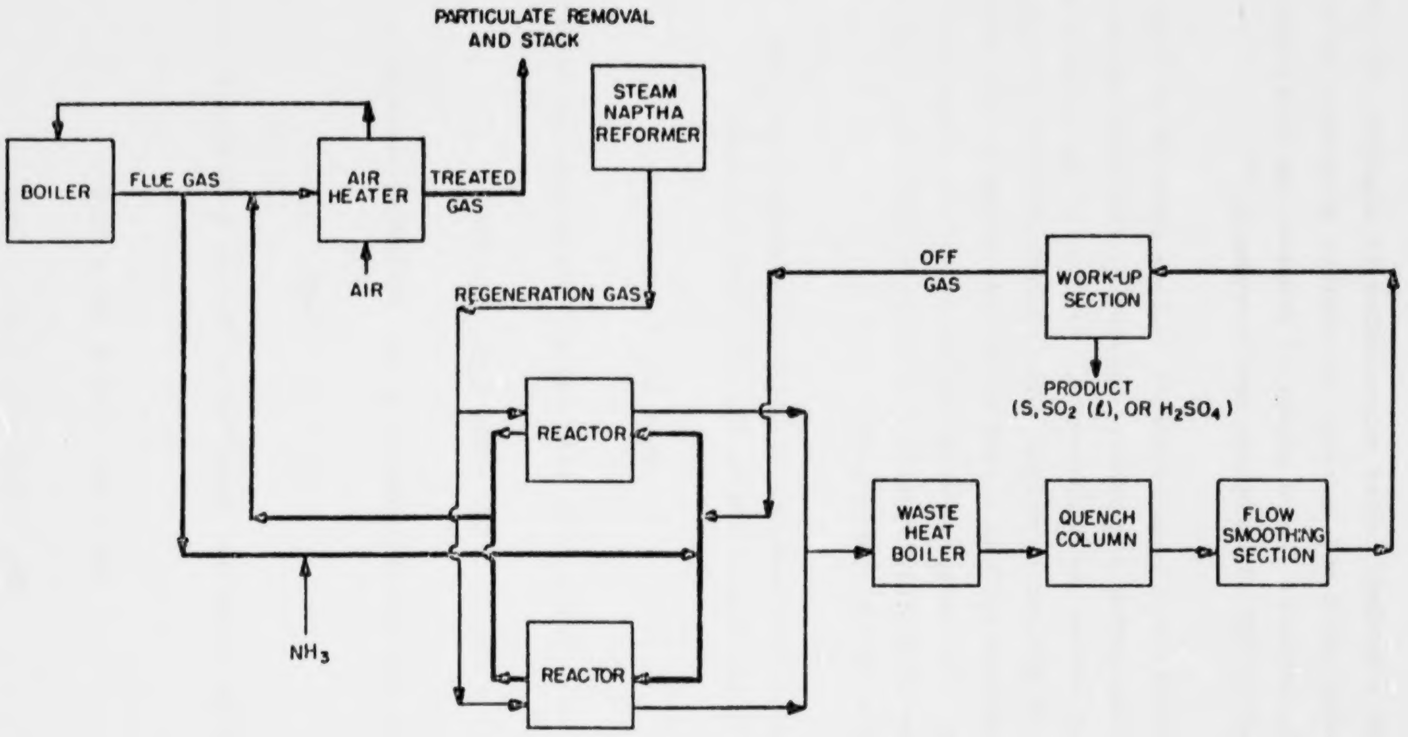

PARTICULATE REMOVAL AND STACK
STEAM NAPTHA REFORMER
BOILER
FLUE GAS
AIR HEATER
TREATED GAS
AIR
REGENERATION GAS
OFF GAS
WORK-UP SECTION
PRODUCT (S, $SO_2$ (ℓ), OR $H_2SO_4$)
REACTOR
REACTOR
WASTE HEAT BOILER
QUENCH COLUMN
FLOW SMOOTHING SECTION
$NH_3$

Figure 2.2.5-2. Flow diagram of the SFGT process.[41]

produced by the reaction. After regeneration is complete, the reactor is again purged with steam and is ready for another acceptance cycle. Regeneration gas can be produced from a number of sources, but steam-naphtha reforming is proposed by UOP as being the most economical.[42]

The regeneration off-gas treatment section consists of flow smoothing equipment and $SO_2$ workup equipment. Typically, the regeneration off-gas is cooled and most of the steam condensed, raising the $SO_2$ concentration from 10 percent to 80 percent by volume. The concentrated $SO_2$ is then compressed into an intermediate holding vessel to provide a smooth flow rate to the workup section. The workup section may be a modified Claus unit which produces an elemental sulfur by-product, a fractionation unit which produces liquid $SO_2$, or a sulfuric acid plant.

Each process step consists of different chemical reactions. The is converted to the oxide form by the following reaction:

$$Cu + \tfrac{1}{2}O_2 \rightarrow CuO \qquad (2\text{-}12)$$

This oxide readily reacts with flue gas $SO_2$ and oxygen, as described by:

$$CuO + \tfrac{1}{2}O_2 + SO_2 \rightarrow CuSO_4 \qquad (2\text{-}13)$$

$SO_3$ in the flue gas is also removed by the following reaction:

$$CuO + SO_3 \rightarrow CuSO_4 \qquad (2\text{-}14)$$

The reaction scheme for reduction of $NO_x$ is described by the following:[12]

$$4NO + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \qquad (2\text{-}1)$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \qquad (2\text{-}2)$$

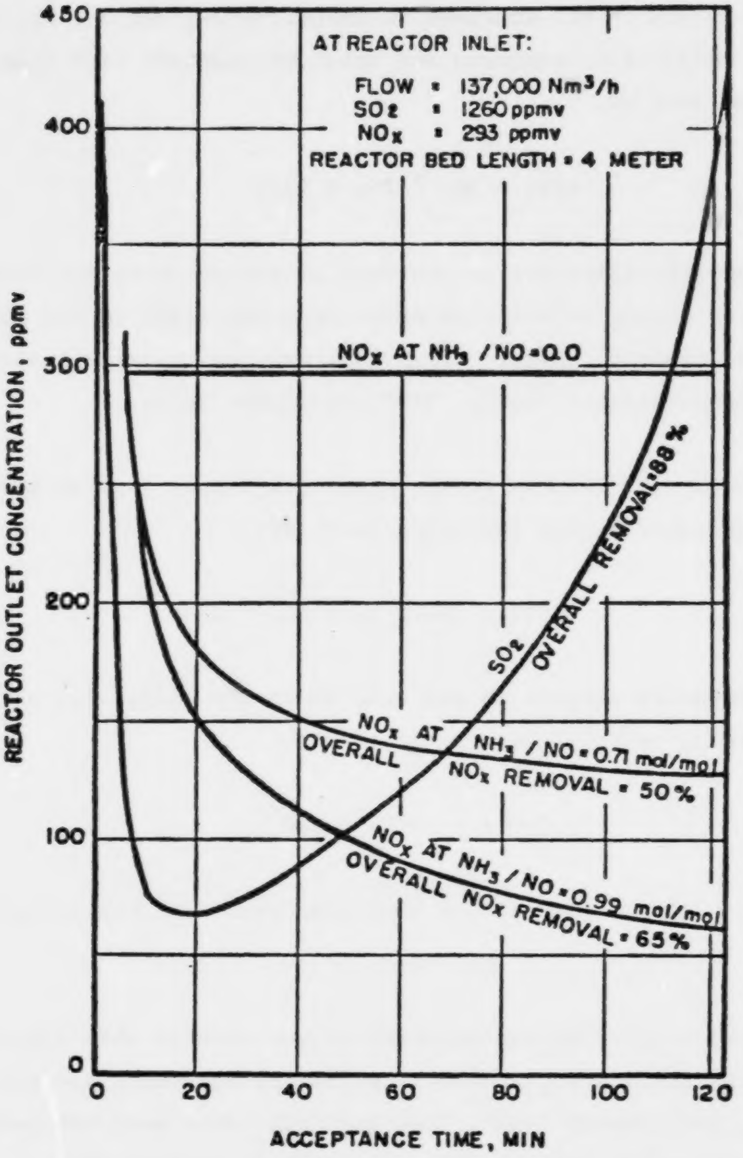


Figure 2.2.5-3. SFGT reactor performance *versus* acceptance time.[44]

Excess ammonia which is not consumed in reactions 2-1 and 2-2 may be catalytically oxidized to nitrogen and water by reaction with flue gas oxygen, as described by:

$$4NH_3 + 3O_2 \rightleftarrows 2N_2 + 6H_2O \qquad (2\text{-}15)$$

Maximum $NO_x$ removal efficiency is achieved at the point of $SO_2$ breakthrough, where conversion of the acceptor material from the oxide to the sulfate form is essentially complete. Figure 2.2.5-3 illustrates reactor outlet $SO_2$ and NO concentrations during a typical SFGT acceptance cycle.

Copper sulfate is reduced to the elemental copper form by reducing gas hydrogen as described by the following reaction:

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O \qquad (2\text{-}16)$$

Any acceptor material present in the reactor as the oxide will also be reduced, according to the following reaction:

$$CuO + H_2 \rightarrow Cu + H_2O \qquad (2\text{-}17)$$

The regeneration step occurs at the same temperature as the acceptance step, 400°C (750°F).

The general reactor design equation is the same as that described in earlier sections for SCR processes. The primary variables are the gas rate, reaction rate, and control level. Reaction rate data have not been released for this process except that the $NO_x$ reduction is first order.

The gas flow rate and control level will determine the reactor size. Increases in either variable will increase the reactor volume. The effect of control level can be seen in Figure 2.2.5-4. It is necessary for the flue gas to enter the reactor at 400°C and therefore it must be taken from

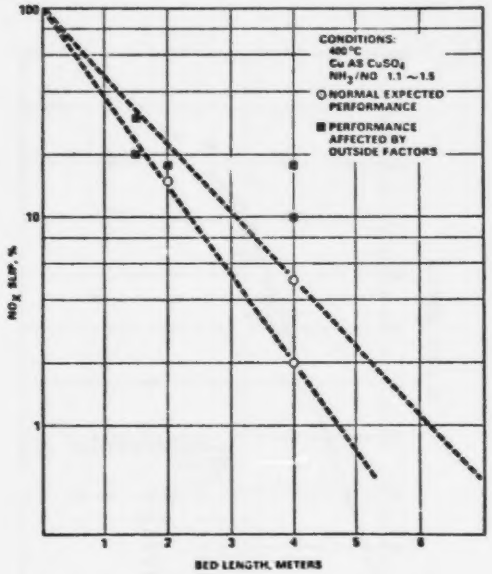


Figure 2.2.5-4. Unconverted $NO_x$ as a function of catalyst bed length.[45]

an appropriate point in the boiler, most likely from between the economizer and air preheater. Alternatively, a cooler gas can be heated to 400°C by an inline heater.

The removal efficiency of $NO_x$ for a given reactor size is determined by the amount of $NH_3$ injected as shown in Figure 2.2.5-5. Since the reaction is first order in $NO_x$, control level is not affected by $NO_x$ concentration.[47] The $SO_2$ control efficiency is primarily a function of the acceptance time of the reactor (Figure 2.2.5-3). Typical ranges of operating variables are shown in Table 2.2.5-1.

Since the SFGT system can handle full particulate loading (≥10 gr/sft$^3$) it is not dependent on any pretreatment facilities. Also, the SFGT system operation is independent of boiler operation. The system fan takes suction from the flue gas duct between the economizer and air preheater and the reactor discharge returns to the boiler flue gas duct just downstream of the

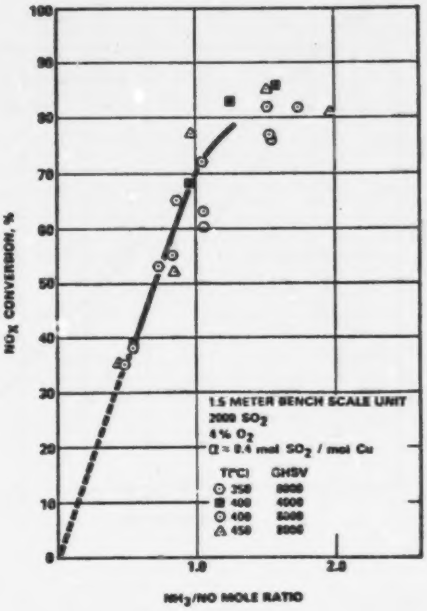


Figure 2.2.5-5. $NO_x$ reduction with $NH_3$ over commercial SFGT acceptor.[46]

TABLE 2.2.5-1. DESIGN AND OPERATING VARIABLES FOR SFGT SYSTEM[48]

| Variable | Typical Range |
|---|---|
| Space Velocity | 5,000 - 8,000 $hr^{-1}$* |
| $NH_3:NO_x$ Mole Ratio | 1.0:1.0 to 1.2:1.0* |
| Flue Gas Temperature | 400°C |
| Pressure Drop | 5-6 in. $H_2O$* |
| Maximum Particulate Loading | ≥23 g/Nm |

*Actual value will depend on required removal level.

suction point, with no valves between the two points. The system fan provides a constant flow rate through the SFGT system. If the boiler flue gas rate is greater than the fan rate, flue gas will bypass the system through the open duct. If the boiler flue gas rate is lower than that of the system fan, treated gas will recycle back to the system fan suction. Recycle of treated gas to the reactor inlet with "open bypass" arrangement presents no operating problems. This is due to the fact that both the level of desulfurization and denitrification are independent of inlet concentrations, and the system does not humidify the flue gas.

Tables 2.2.5-2 and 2.2.5-3 present test and commercial applications of the SFGT process. The development history of the process can also be seen in these tables.

In the U.S., from 1974 to 1976 a pilot scale unit at Tampa Electric Company (TECO) was operated using flue gas from a coal-fired boiler. Testing was for $SO_2$ removal only, $NO_x$ control was not attempted during this period. The process developer is currently modifying the TECO pilot unit to accommodate 7 meters of bed height, up from the previous maximum of 5 meters. This should permit simultaneous removal of $NO_x$ and $SO_x$ to the 90 percent level. Also, provisions are being made for injection of a $CO/CO_2$ gas mixture into the regeneration gas in order to simulate medium-Btu gas from a coal gasifier.

The costs for an industrial size boiler have not been estimated. However, costs for a 500 MW utility boiler application are available and are shown in Tables 2.2.5-4, 5, and 6. Also shown are the estimated energy and raw material requirements.

TABLE 2.2.5-2. SFGT PROCESS, PILOT AND DEMONSTRATION UNITS

| Location/ Company | Designed By | Fuel/ Application | Size, $Nm^3$/hr | Type of Operation | Dates | Comments |
|---|---|---|---|---|---|---|
| Shell Ref. at Pernis | Shell | Residual Fuel Oil-Proc. Heater | 600-1000 | $SO_x$-only | 1967-1972 | $SO_x$ reduction - approx. 90% |
| Rotterdam Utility | Shell | Coal-Steam Boiler | - | Heavy Fly Ash Loading | 1971 | Particulate matter - loadings to 20 $Gr/Nm^3$ |
| Tampa Elec. Big Bend | UOP | Coal-Wet-Bottom Utility Boiler | 1200-2000 | $SO_x$-only;<br>$SO_x$-$NO_x$ Simultaneous | 1974-1976<br>1979- | $SO_x$ - 90%;<br>$SO_x$-$NO_x$ - 90/90% fly ash to 25 $Gr/Nm^3$ |
| JGC Yokohama | JGC* | Fuel Oil | 250-700 | $NO_x$-only | 1974- | $NO_x$ reduction - 90-99% |
| Nippon Steel | JGC | Sintering Furnace | 2000 | $NO_x$-only | 1976-1978 | $NO_x$ reduction - 90-97% |
| -- | JGC | Coke Oven Gas | 400 | $NO_x$-only | 1976-1977 | $NO_x$ reduction - 90%; special low temp. cat. evaluation |

*JGC Corporation, licensing agent in Japan.

TABLE 2.2.5-3. SFGT PROCESS, COMMERCIAL UNITS

| Unit | Designed By | Fuel/ Application | Size, $Nm^3/hr$ | Type of Operation | Dates | Comments |
|---|---|---|---|---|---|---|
| SYS* Yokkaichi | Shell | Residual Fuel Oil-Ref. Boiler | 125,000 | $SO_x$-only;<br>$NO_x$-$SO_x$ Simultaneous | 1973-1975<br>1975- | $SO_x$ reduction - 90%;<br>Simultaneous - 90/50% |
| Kashima Oil Co. Ltd. | JGC | Fuel Oil-Process Unit Heater | 50,000 | $NO_x$-only | 1975- | 95-98% |
| Fuji Oil Co. Ltd. | JGC | CO Boiler | 70,000 | $NO_x$-only | 1976- | 93-96% |
| Nippon Steel Corp. | JGC | Sintering Furnace | 150,000 | $NO_x$-only | 1978- | ∿95% (low temp. catalyst) |

*Showa Yokkaichi Sekiyu

## TABLE 2.2.5-4. ECONOMICS OF SFGT SYSTEM[49]

| | BASIS: |
|---|---|
| Incorporated Units: | Steam-Naphtha Reformer<br>SFGD Reactor Section<br>Compressor/Gasholder Flow Smooth Section<br>Modified Claus Unit |
| Power Plant Size | 500 MW |
| Fuel | Coal |
| S-Content, Wt-% | |
| Case 1 | 3.5 |
| Case 2 | 2.5 |
| Case 3 | 0.8 |
| HHV | 10,500 Btu/lb |
| Heat Rate | 9,000 Btu/kWh |
| Excess Air | 20% |
| Air Preheater Leakage | 13% |
| Flue Gas Rate | 1,582,000 $Nm^3$/h (983,000 SCFM) |
| $SO_2$ Content, ppmv | |
| Case 1 | 2,580 |
| Case 2 | 1,850 |
| Case 3 | 590 |
| Mid-1977, Gulf Coast Location | |
| Load Factor | 7,000 h/a |
| Capital Charges | 15%/a |
| Cost of: | |
| Naphtha | $0.35/gal |
| Steam (40 psi, SAT.) | $1.50/M lb |
| Electricity | $0.018/kWh |
| Labor | $10.00/hr |
| Heat Credits | $2.50/MMBtu |
| Sulfur | $45.00/ton |

TABLE 2.2.5-5. ECONOMICS OF SFGT SYSTEM ESTIMATED CHEMICALS AND UTILITY REQUIREMENTS[50]

| | | SFGD Section | Flow Smooth Section | Mod. Claus Section | Reformer Section | Total |
|---|---|---|---|---|---|---|
| Case 1 | | | | | | |
| Electricity | kW | 5,770 | 850 | 115 | 480 | 7215 |
| Steam** | kmol/h | 1,820 | -380* | -740* | -600* | 100 |
| Naphtha*** | Gcal/h | | | | 90.92 | 90.92 |
| Heat Credits | Gcal/h | | | | | 42.53 |
| S° Produced | kg/h | | | 5250 | | 5250 |
| Case 2 | | | | | | |
| Electricity | kW | 5,800 | 570 | 82 | 300 | 6782 |
| Steam** | kmol/h | 1,300 | -270* | -530* | -415* | 85 |
| Naphtha*** | Gcal/h | | | | 62.75 | 62.75 |
| Heat Credits | Gcal/h | | | | | 32.48 |
| S° Produced | kg/h | | | 3760 | | 3760 |
| Case 3 | | | | | | |
| Electricity | kW | 5,120 | 180 | 30 | 110 | 5440 |
| Steam** | kmol/h | 480 | -95* | -170* | -140* | 75 |
| Naphtha*** | Gcal/h | | | | 21.01 | 21.01 |
| Heat Credits | Gcal/h | | | | | 18.46 |
| S° Produced | kg/h | | | 1200 | | 1200 |

*Produced

**40 psig, Saturated

***5.175 MMBtu/Bbl produces 11,500 SCF Hydrogen/Bbl

## TABLE 2.2.5-6. ECONOMICS OF SFGT SYSTEM ESTIMATED CAPITAL AND OPERATING COST[51]

| | Case 1 | Case 2 | Case 3 |
|---|---|---|---|
| EEC. (MM$) | | | |
| SFGD Reactor Section | 28.95 | 28.53 | 22.94 |
| Compressor/Gasholder | 7.82 | 6.10 | 2.65 |
| Modified Claus | 2.76 | 2.26 | 1.14 |
| Steam-Naphtha Reformer | 8.81 | 7.14 | 4.17 |
| Estimated Annual Revenue Requirements (M$/a) | | | |
| Capital Charges | 7251 | 6604 | 4634 |
| Maintenance | 967 | 881 | 618 |
| Labor | 123 | 123 | 123 |
| Acceptor | 1479 | 1053 | 411 |
| Electricity | 909 | 855 | 685 |
| Steam | 42 | 35 | 31 |
| Naphtha | 7174 | 4951 | 1658 |
| Heat Credits | -2977 | -2273 | -1292 |
| Sulfur Credits | -1570 | -1126 | -359 |
| Capital Cost, Operating Cost, Energy Requirement | | | |
| Capital Cost, $/kW | 97 | 88 | 62 |
| Operating Cost, ¢/kWh | 0.38 | 0.32 | 0.19 |
| Energy Requirement, Btu/kWh* | 525 | 371 | 124 |

*Defined as the sum of:

| | |
|---|---|
| Electricity at | 9000 Btu/kWh |
| Steam at | 40000 Btu/kmol |
| Naphtha at | 4 Btu/kcal |
| Heat Credits at | 4 Btu/kcal |

### 2.2.5.2 System Performance--

$NO_x$ control by the SFGT process is shown graphically in Figure 2.2.5-5. As can be seen, at a space velocity of 8000 $hr^{-1}$, $NO_x$ control of >80 percent can be achieved. Figure 2.2.5-4 indicates that the process developers feel the process to be capable of $NO_x$ control levels of >90 percent.

Several different test series were conducted using the TECO pilot plant and the operating conditions for these tests are shown in Table 2.2.5-7. The $SO_2$ removal efficiency for several of these runs is shown in Figure 2.2.5-6 plotted against the number of cycles, which can be converted to time. No data of this type are available for $NO_x$ control using coal-fired flue gas, however, these data should be available in about one year.

As mentioned earlier, the system is not impacted by changes in the boiler gas rate or particulate concentrations. Changes in the $NO_x$ concentration due to boiler load changes can be compensated for by a conventional control system used in conjunction with the $NH_3$ injection equipment. This control system will be developed during the upcoming pilot tests at the TECO pilot plant.

## 2.2.6 Adsorption

### 2.2.6.1 System Description--

The adsorption process removes $NO_x$ and $SO_2$ from flue gas by adsorbing them onto a special activated char. Adsorbed $NO_x$ is reduced to $N_2$ while $SO_2$ is reduced and condensed to an elemental S by-product.

A process flow diagram is shown in Figure 2.2.6-1. Flue gas is taken from the boiler air preheater and passed through a particulate removal device to prevent blinding of the adsorption bed. The flue gas then enters the adsorber, a vertical column with parallel louver beds containing the char in

TABLE 2.2.5-7. SUMMARY OF BASE OPERATING CONDITIONS ON THE SFGT PILOT PLANT AT TECO[52]

| Run No. | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| Duration, Months | 5 | 2½ | 1½ | 2 | 5 | 3 |
| Cycles | 2488 | 1520 | 1292 | 1412 | 4328 | 2210 |
| Cumulative Cycles | 2488 | 4008 | 5300 | 6712 | 11040 | 13250 |
| Flow Rate, SCFM | 1090 | 1090 | 1090 | 1090/1420 | 1090 | 1090 |
| Acc. Time, Min. | 20 | 20 | 20 | 20 | 20 | 20 |
| Reg. Time, Min. | 20 | 20 | 20 | 20 | 20 | 20 |
| Flue Gas Source* | 1 | 1 | 2 | 1 | 1 | 3 |
| Eff. SOR |  | 92 | 95 | 95 | 95 | 93 |
| Eff. EOR |  | 82 | 95 | 80 | 92 | 93 |

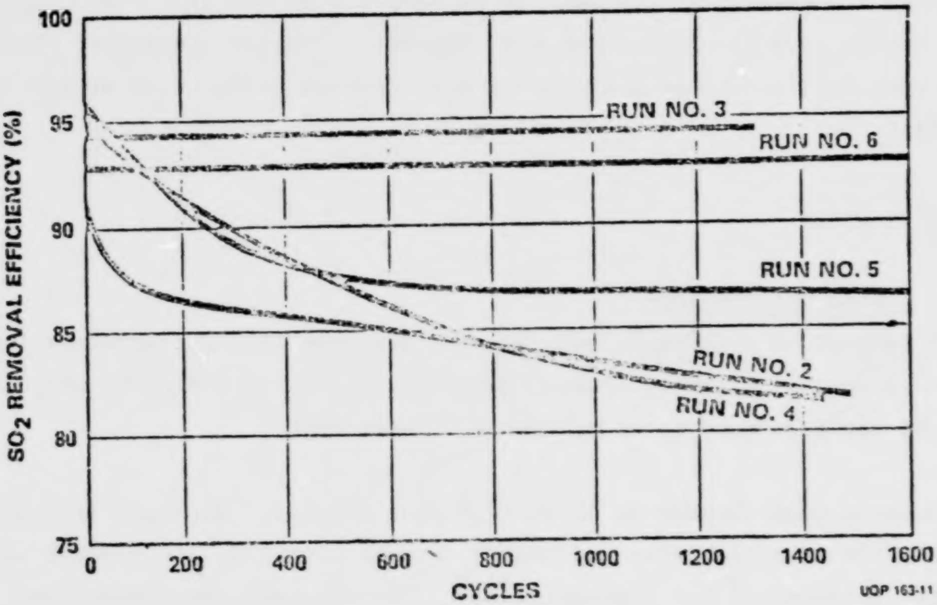


Figure 2.2.5-6. $SO_2$ removal efficiency *vs.* cycles.[53]

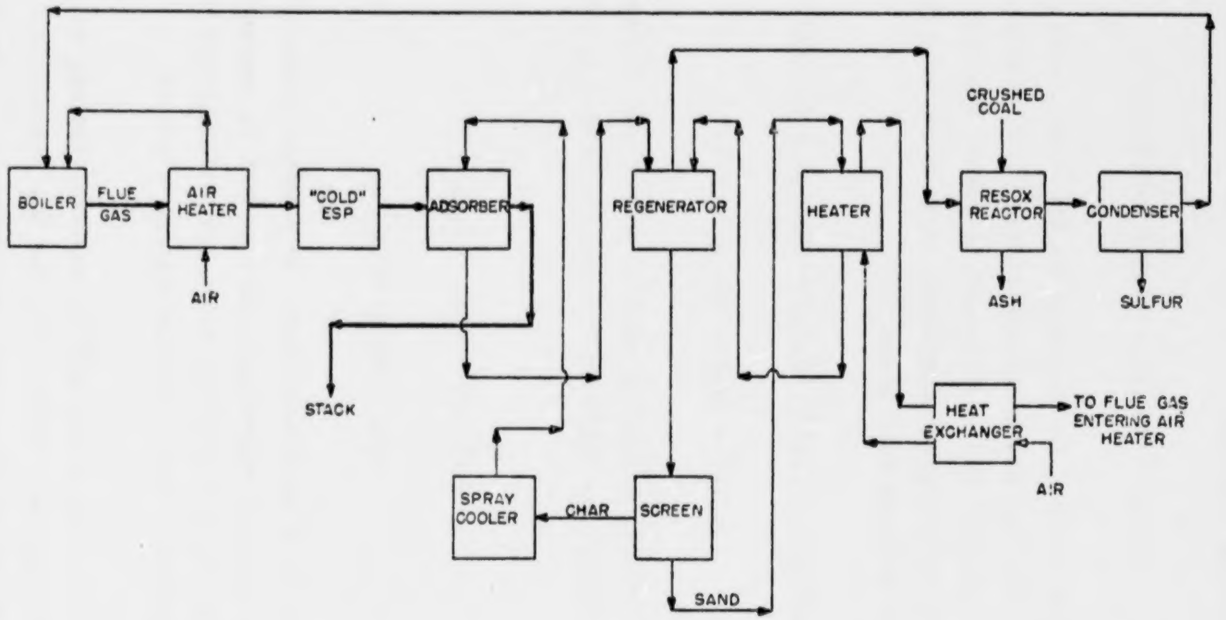

BOILER
FLUE GAS
AIR HEATER
AIR
"COLD" ESP
ADSORBER
STACK
REGENERATOR
HEATER
CRUSHED COAL
RESOX REACTOR
ASH
CONDENSER
SULFUR
HEAT EXCHANGER
TO FLUE GAS ENTERING AIR HEATER
AIR
SPRAY COOLER
CHAR
SCREEN
SAND

Figure 2.2.6-1. Flow diagram of Foster Wheeler-Bergbau Forschung Dry Adsorption Process.[54]

pellet form. $NO_x$ and $SO_2$ are adsorbed on the char which slowly moves downward through the bed. The $NO_x$ adsorption mechanism is unknown but $SO_2$ undergoes the following reaction.

$$SO_2(g) + H_2O(g) + \tfrac{1}{2}O_2(g) \rightarrow H_2SO_4(l) \qquad (2\text{-}18)$$

The reaction product is held in the pores of the char pellets. The flue gas exits the adsorber and passes to the stack. The saturated char leaves the bottom of the adsorber and is screened to remove any fly ash deposits. It is then conveyed to a regenerator where it is mixed with hot sand (650°C) and the following reactions take place.[55,56]

$$2H_2SO_4(l) + C(s) \rightarrow CO_2(g) + 2H_2O(g) + 2SO_2(g) \qquad (2\text{-}19)$$

$$2NO(g) + C(s) \rightarrow CO_2(g) + N_2(g) \qquad (2\text{-}20)$$

This $SO_2$-rich gas product stream is sent to an off-gas treatment reactor containing hot, crushed coal (650-820°C) and the following reactions take place.[56]

$$SO_2(g) \rightarrow S(g) + O_2(g) \qquad (2\text{-}21)$$

$$C(s) + O_2(g) \rightarrow CO_2(g) \qquad (2\text{-}22)$$

The gas then passes to a condenser where the S vapor forms molten S. The char/sand mixture from the regenerator is screened to separate the two solids. The char is recycled to the adsorber via a spray cooler and the sand is recycled to the regenerator after passing through a heater.

This process operates at 120-150°C, however, typical values for other operating variables were not found. $NO_x$ and $SO_2$ control levels were reported to be 40-60 percent and 80-95 percent, respectively.[57] The economics of the process vary with the fuel sulfur level. For fuel sulfur levels of 0.9-4.3

percent, the capital costs range from \$40-90/kW and the operating costs range from 1.0-2.3 mills/kWh.[58] The costs were based on applying the process to a utility boiler of >200 MW capacity.

Presently, the adsorption process is in the prototype unit stage of development. The one reported process developer in the field, Foster Wheeler-Bergbau Forschung has a 20 MW prototype unit and several small pilot plants treating coal-fired flue gas.

### 2.2.6.2 System Performance--

Tests have shown the adsorption process to be primarily a $SO_2$ reduction process as $NO_x$ removal efficiency averages 40-60 percent while $SO_2$ removal had a range of 80-95 percent.[59]

The primary drawback of this process, besides the low $NO_x$ removal level, is its complexity: numerous process steps involving hot solids handling. Solids flow can be difficult to control and high maintenance requirements could be expected. The vendor has reported several mechanical problems during testing which included control of adsorber-bed levels, poor char distribution, char-sand separation, hot sand conveying, and char cooling and feed. Some corrosion-resistant material is needed in the high temperature zones of the process. The ash waste stream from the off-gas treatment reactor appears to be the sole secondary pollutant associated with the process. The overall complexity and low $NO_x$ removal of the process present definite technical disadvantages.

## 2.2.7 Electron Beam Radiation

### 2.2.7.1 System Description--

This dry process utilizes an electron beam to bombard the flue gas, removing $NO_x$ and $SO_2$ in the process. A block flow diagram for the process is shown in Figure 2.2.7-1.

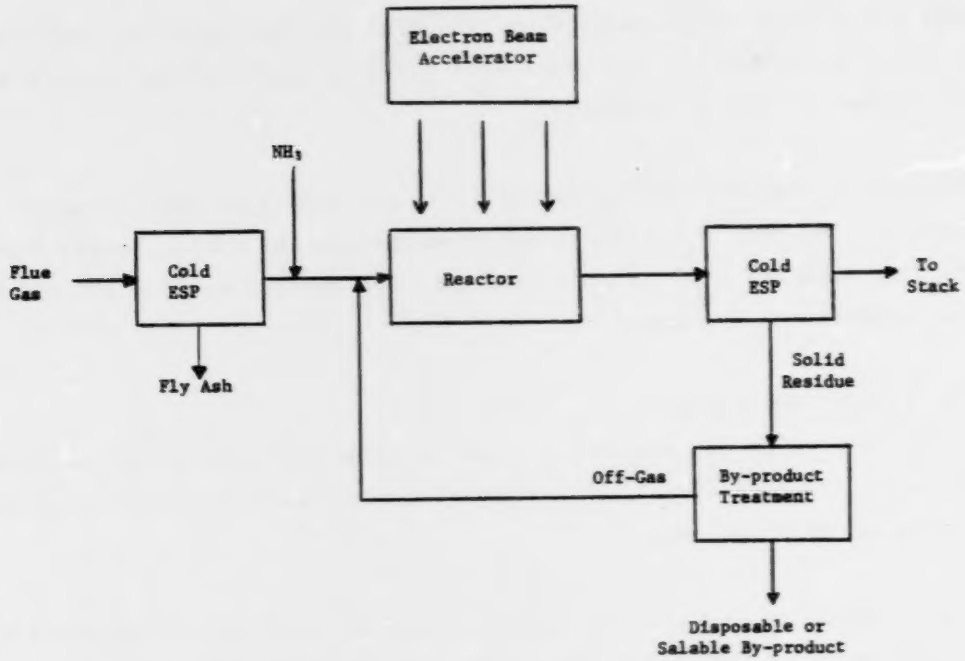

Electron Beam Accelerator
NH$_3$
Flue Gas
Cold ESP
Reactor
Cold ESP
To Stack
Fly Ash
Solid Residue
Off-Gas
By-product Treatment
Disposable or Salable By-product

Figure 2.2.7-1. Process flow diagram for Ebara-JAERI electron beam process.[60]

Flue gas is taken from the boiler air preheater and passed through a cold ESP to remove particulates. After a small amount of ammonia is added, the gas enters a reactor where it is bombarded with an electron beam. (The penetration of the gas stream by the beam will require a unique discharge pattern or other special design considerations.) A powder containing both ammonium nitrate and sulfate is generated by an unknown reaction mechanism. The gas then exits the reactor, passes through a second ESP to remove the solid by-product, and is sent to the stack. The by-product treatment system is still being developed. Various methods investigated include thermal decomposition in the presence of an inert gas, steam roasting with CaO, or steam roasting with $H_2O$. The by-product may eventually be useful as a fertilizer.[61]

The key subsystem of this process is the electron beam accelerator. Control of this unit's power supply is based upon inlet composition, flow rate, and temperature of the flue gas.

Some of the important variables and typical ranges are listed in Table 2.2.7-1.

TABLE 2.2.7-1. SYSTEM VARIABLES[62]

| | Typical Value |
|---|---|
| Temperature | ~100°C |
| Reactor residence time | 1-20 sec |
| Radiation rate | $10^5$-$10^6$ rad*/sec |
| Total radiation absorbed | 1-3 Mrad* |

*Rad is the radiation dose absorbed
1 rad = .01 J/Kg

The operating cost with $NO_x$ removal only (low sulfur coals) is lower due to lower radiation levels, but the capital cost would be just as high as for simultaneous $NO_x/SO_x$ removal. Capital costs are quite high for this process as the 2 ESP's and the accelerator are expensive. The cost for a 1000 $Nm^3$/hr test unit are reported tc be $1000/kW, however, the cost of a full scale system is expected to be lower. Operating costs are not available.

No coal-fired tests have been performed at this time. The Ebara Manufacturing Company in conjunction with Japan Atomic Energy Research Institute (JAERI) has operated a 1000 $Nm^3$/hr pilot plant treating flue gas from an oil-fired boiler. In 1976, a 3000 $Nm^3$/hr pilot plant began treating off-gas from an iron ore sintering furnace at Nippon Steel. By-product treatment technology needs to be more fully developed before this process can be applied commercially.

In the U.S., the Department of Energy (DOE) is funding development of an electron beam process offered by Research-Cottrell. Pilot unit tests with flue gas are scheduled, however, the details of the program are not yet available.

2.2.7.2 System Performance--

No coal-fired testing has been done.

A summary of the oil-fired pilot tests is shown in Figure 2.2.7-2.

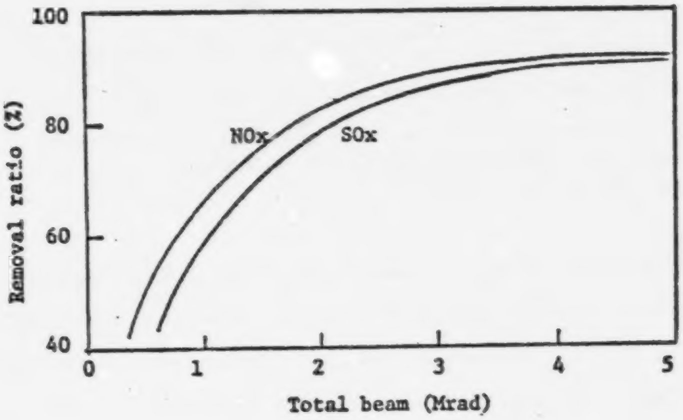


Figure 2.2.7-2. Oil-fired pilot plant test results.[64]

One can see that $NO_x/SO_2$ removal drops off drastically at a total radiation dose below 1 Mrad while the maximum removal is obtained at about 3 Mrad. The removal efficiencies decrease as the concentrations of $NO_x$ and $SO_2$ increase as can be seen in Figure 2.2.7-3.

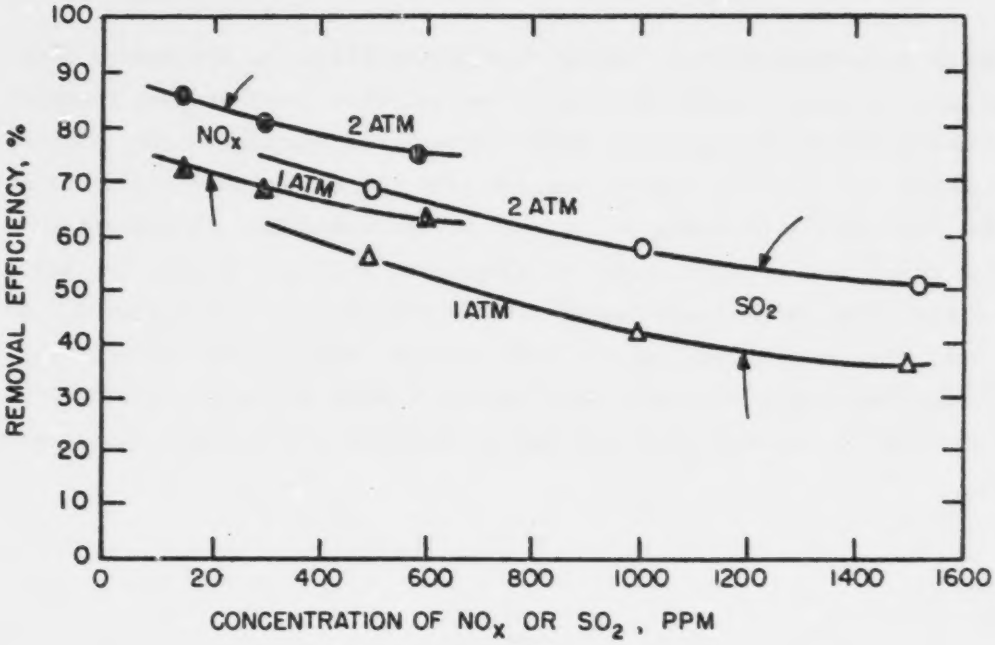


Figure 2.2.7-3. Effect of pollutant concentration on removal efficiency.[65]

## 2.2.8 Absorption-Reduction

### 2.2.8.1 System Description--

Absorption-reduction processes simultaneously remove $NO_x$ and $SO_2$ from flue gas by absorbing them into a scrubbing solution. The processes are based on the use of chelating compounds, such as ethylenediamine tetraacetic acid (EDTA) complexed with iron, to "catalyze" the absorption of $NO_x$. Most process vendors prefer a perforated-plate type of gas-liquid contactor. The advantages of a perforated-plate absorber over a packed bed absorber include easier cleaning when solids are present, wider operating ranges, and more economical handling of high liquid rates.[66] An example of a perforated plate absorber is shown in Figure 2.2.8-1. The most common design of a perforated plate is one that employs liquid crossflow over the face of the plate with the gas passing upward through the plate perforations. A schematic of the operation of a crossflow perforated plate is shown in Figure 2.2.8-2. The liquid is prevented from flowing through the plates by the upward flow of the gas. However, during periods of low gas flow (such as load changes on industrial boilers) liquid can drain through the openings in the plates. This reduces the liquid's time of contact with the gas on each plate and may decrease the overall operating efficiency of the absorber. To prevent this problem, there are two other types of dispersers utilized besides the basic sieve-plate: the valve-plate and the bubble cap, depicted in Figure 2.2.8-3. As the gas flow lowers, the valve or cap settles, sealing off the perforation so liquid cannot drain through. This design feature allows the perforated plate absorber to operate more efficiently at widely fluctuating gas rates.

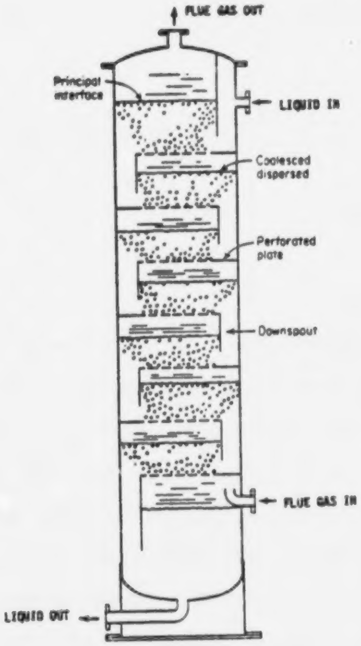

FLUE GAS OUT
Principal interface
LIQUID IN
Coalesced dispersed
Perforated plate
Downspout
FLUE GAS IN
LIQUID OUT

Figure 2.2.8-1. Perforated plate absorber option for Absorption-Reduction Processes.[28]

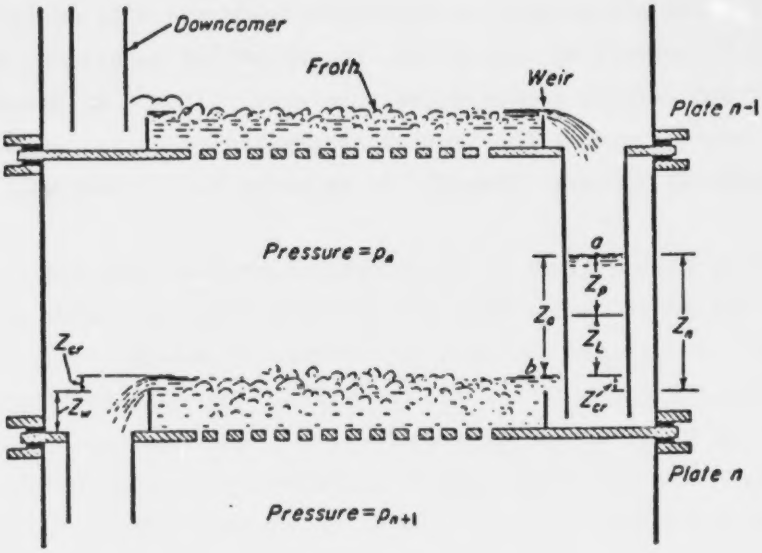
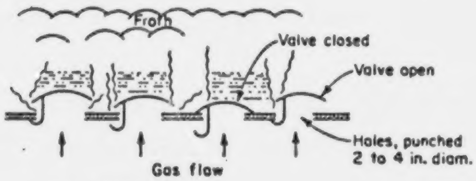
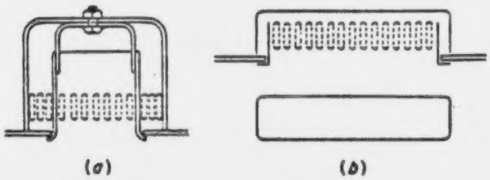

Downcomer
Froth
Weir
Plate n-1
Pressure = $p_n$
a
$Z_p$
$Z_a$
$Z_L$
$Z_n$
b
$Z_{cr}$
$Z_{cr}$
$Z_w$
Plate n
Pressure = $p_{n+1}$

Figure 2.2.8-2. Normal operation of sieve plate. $Z_a$, height of station a above datum. $Z_{cr}$, weir crest. $Z_L$, liquid-friction head. $Z_p$, pressure head across plate. $Z_n$, net head in down pipe. $Z_w$, weir height.[67]


Froth
Valve closed
Valve open
Holes, punched
2 to 4 in. diam.
Gas flow

**Valve-plate dispersers.**



**(a) Circular or bell cap. (b) Tunnel cap.**

Bubble cap dispersers

Figure 2.2.8-3. Other gas dispersers.[68]

While most all absorption-reduction processes utilize ferrous chelating compounds to enhance NO absorption, the scrubbing solutions, the by-product treatment and sorbent regeneration chemistry differ from process to process. For this reason, one of the simpler absorption-reduction processes, that of Kureha Chemical Industry Company, is examined here in detail.

A block flow diagram of the Kureha absorption-reduction process is shown in Figure 2.2.8-4. Flue gas is taken from the boiler after the air preheater. It passes through a prescrubber to adiabatically cool the gas and remove both particulates and chlorides. The flue gas then enters the distributing space at the bottom of the $NO_x/SO_2$ absorber, below the plates or packing. The gas flows upward, countercurrent to a sodium acetate ($CH_3COONa$) scrubbing solution (∿60°C) containing ferrous iron and EDTA and a few seed crystals of gypsum (to prevent scaling). Most of the $SO_2$ is rapidly absorbed at the bottom of the absorber according to the following reactions.[71]

$$SO_2(g) \rightarrow SO_2(aq) \tag{2-23}$$

$$SO_2(aq) + 2CH_3COONa(aq) + H_2O \rightarrow Na_2SO_3(aq) + 2CH_3COOH(aq) \tag{2-24}$$

The $NO_x$ (which consists mainly of NO) is relatively insoluble; therefore, it is absorbed gradually over the length of the column. The ferrous chelating compounds effect on NO absorption is described in Figure 2.2.8-5. The $NO_x$ is absorbed and undergoes the following reactions.[73]

$$NO(g) \rightarrow NO(aq) \tag{2-6}$$

$$2NO_2(g) \rightarrow N_2O_4(aq) \tag{2-25}$$

$$N_2O_4(g) \rightarrow N_2O_4(aq) \tag{2-9}$$

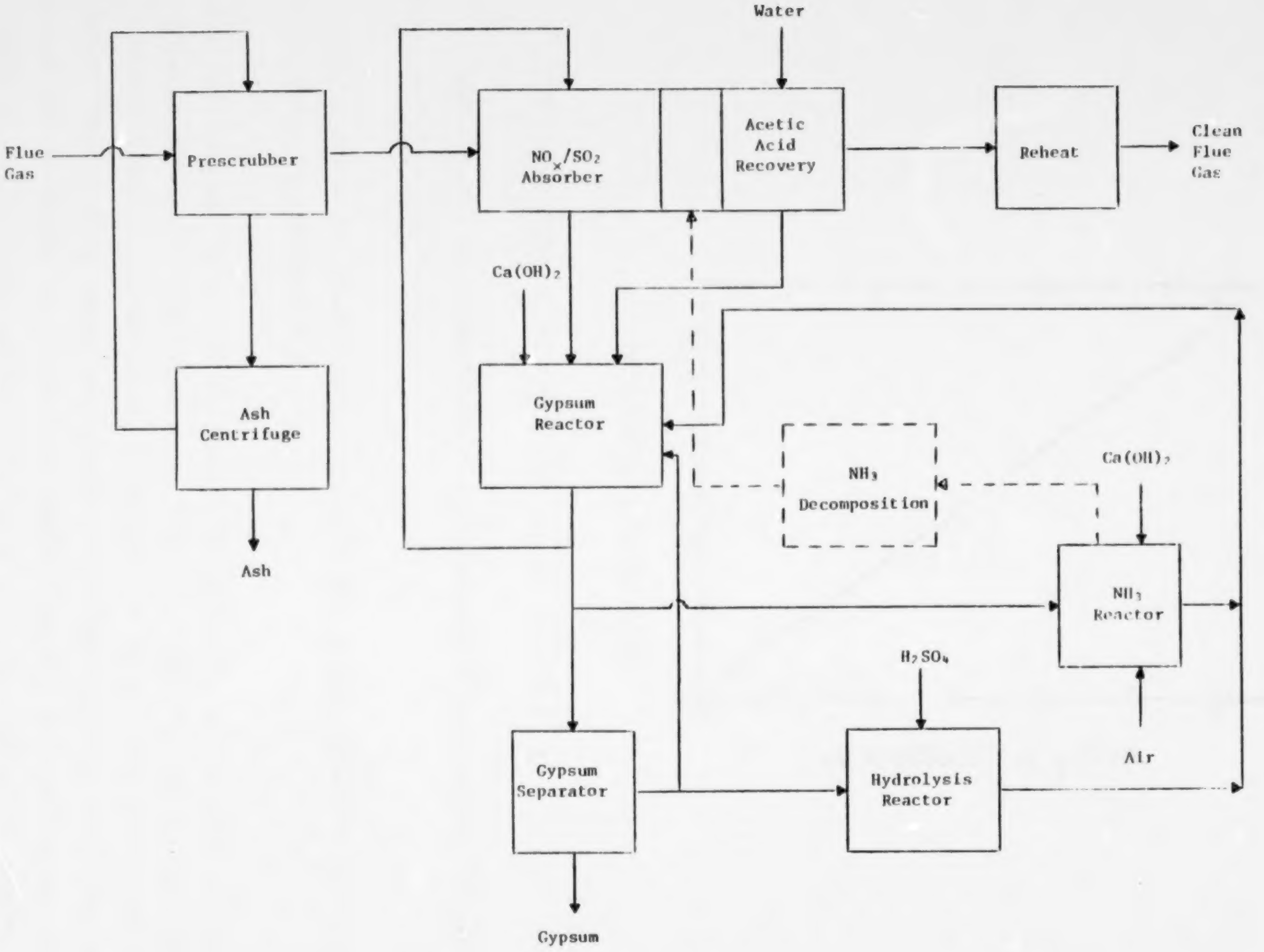

Water
Flue Gas
Prescrubber
NO_x/SO_2 Absorber
Acetic Acid Recovery
Reheat
Clean Flue Gas
Ca(OH)_2
Gypsum Reactor
Ash Centrifuge
Ash
NH_3 Decomposition
Ca(OH)_2
NH_3 Reactor
H_2SO_4
Air
Gypsum Separator
Hydrolysis Reactor
Gypsum

Figure 2.2.8-4. Process flow diagram of Kureha absorption-reduction process.[69,70]

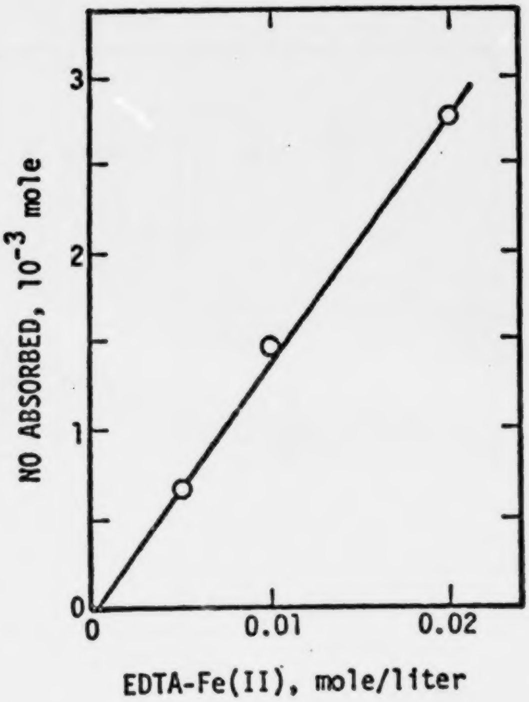


Figure 2.2.8-5. EDTA-Fe(II) concentration and NO absorption at 50°C.[72]

$$2NO(aq) + 5Na_2SO_3(aq) + 4CH_3COOH(aq) \rightarrow 2NH(SO_3Na)_2(aq) + Na_2SO_4(aq) + 4CH_3COONa(aq) + H_2O \quad (2\text{-}26)$$

$$2N_2O_4(aq) + 7Na_2SO_3(aq) + 4CH_3COOH(aq) \rightarrow 2NH(SO_3Na)_2(aq) + 3Na_2SO_4(aq) + 4CH_3COONa(aq) + H_2O \quad (2\text{-}27)$$

Some of the acetic acid ($CH_3COOH$) formed at the bottom of the absorber via reaction (2-24) is vaporized. It must be captured and is done so by water scrubbing at the very top of the absorber. From the top of the absorber column the clean flue gas passes to a heater for plume buoyancy and is then sent to the stack. The liquid effluent drops from the bottom of the absorber to a gypsum, $CaSO_4 \cdot 2H_2O$, production reactor. Here, the solution is mixed with with the purge stream from the acetic acid recovery section and a lime slurry stream. The lime, $Ca(OH)_2$, treatment involves the following reactions.[74]

$$2CH_3COOH(aq) + Ca(OH)_2(aq) \rightarrow (CH_3COO)_2Ca(aq) + 2H_2O \qquad (2\text{-}28)$$

$$(CH_3COO)_2Ca(aq) + Na_2SO_4(aq) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s)\downarrow + 2CH_3COONa(aq) \quad (2\text{-}29)$$

The gypsum formed by reaction 2-29 is centrifuged. Most of the liquor discharged is returned to the gypsum reactor and on to the absorber. The remaining liquor is sent to a reactor where sulfuric acid ($H_2SO_4$) is added to hydrolyze the imidodisulfonate, $NH(SO_3Na)_2$, by the following reaction.[75]

$$NH(SO_3Na)_2(aq) + 2H_2O \xrightarrow[140°C]{H^+} NH_4HSO_4(aq) + Na_2SO_4(aq) \qquad (2\text{-}30)$$

The effluent from this reactor is then recycled to the gypsum production reactor. A small purge stream is taken from the gypsum reactor to another reactor where the ammonium bisulfate ($NH_4HSO_4$) formed in the hydrolysis reaction is treated with lime to yield gypsum and $NH_3$ off-gas by the following reaction.[76]

$$NH_4HSO_4(aq) + Ca(OH)_2(s) \rightarrow CaSO_4 \cdot 2H_2O(s)\downarrow + NH_3(g)\uparrow \qquad (2\text{-}31)$$

The gaseous ammonia is stripped from the solution by an air stream. If no use for the ammonia can be found, the gas mixture is sent to a catalytic reactor where ammonia is oxidized by the following reaction.[77]

$$4NH_3(g) + 3O_2(g) \xrightarrow[350°C]{catalyst} 2N_2(g) + 6H_2O(g) \qquad (2\text{-}32)$$

The product stream is then sent to the deacetating section of the absorber column.

The fundamental design equation used for gas absorption column design is[32]

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \tag{2-11}$$

where $y$ = bulk $NO_x$ concentration (mole fraction of gas phase at any given point in column

$y-y^*$ = overall driving force for absorption ($y^*$ being the $NO_x$ concentration of a gas in equilibrium with given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/($ft^2$)(hr)(mole fraction)

$a$ = area of gas-liquid interface per unit packed volume, $ft^2/ft^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/($ft^2$)(hr)

$Z$ = length of packed section of column, ft

In a column containing a given plate or packing configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity, obviously, must be lower than the flooding velocity. How much lower is a design decision. Often it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[33] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter consideration is the reason why a chelating compound is used in absorption-reduction processes to aid in $NO_x$ absorption. Table 2.2.8-1 presents the effects of boiler/flue gas variables on the design of absorption-reduction systems. Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load change. The column size and the liquid flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

TABLE 2.2.8-1. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Requires $SO_2$:$NO_x$ mole ratio of at least 3-5[9] (depending on process) for absorption-reduction to be effective. |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased catalyst concentration |

The process vendors have not released much information on the operating conditions of these processes. This is primarily due to the competitive status of these similar processes at this early stage of development. Typical values for some of the process variables are shown in Table 2.2.8-2.

TABLE 2.2.8-2. TYPICAL VALUES FOR PROCESS VARIABLES OF ABSORPTION-REDUCTION PROCESSES[78]

| Variable | Range |
|---|---|
| Liquid/Gas ratio, $1/Nm^3$ | 10-30 |
| $SO_x/NO_x$ mole ratio | 2.5-3.0 |
| Superficial Gas Velocity, m/s | 1-3 |

Cost estimates for this type of process cover a large range, presumably due to the differences in sorbent regeneration technique. Capital costs for utility applications are reported to range from $65-127/kW and operating costs from 4.8-7.4 mills/kWh.[79]

Presently, absorption-reduction processes are in the pilot-unit stage of development. Table 2.2.8-3 presents a list of absorption-reduction process vendors and the status of development of their projects. One can see from the table that only one coal-fired flue gas test has been performed.

TABLE 2.2.8-3. PROCESS VENDORS OF ABSORPTION-REDUCTION PROCESSES[80]

| Vendor | Status of Development |
|---|---|
| Asahi | 1974: 600 $Nm^3$/hr flue gas from residual oil-fired boiler (1000 hours continuous). |
| Chisso | 1975: 300 $Nm^3$/hr flue gas from oil-fired boiler (335 hours continuous) |
| Kureba | 1976: 5000 $Nm^3$/hr flue gas from heavy oil-fired boiler (3000 hours continuous) |
| Mitsui Engineering and Shipbuilding | 1974: 150 $Nm^3$/hr flue gas from oil-fired boiler |
| Pittsburgh Environmental | 1976: 3000 $Nm^3$/hr flue gas from coal-fired boiler (52 hours continuous, absorption section only) |

### 2.2.8.2 System Performance--

The one coal-fired test showed 60-70 percent $NO_x$ and 90 percent $SO_2$ reductions are possible.[11] The longest continuous operation was for 52 hours and the absorption section was the only part of the process tested. Pilot-plant testing was discontinued after two months. Plans are being made for further coal-fired pilot tests on the integrated system.

Absorption-reduction processes are readily applicable only to high sulfur coals as a $SO_2:NO_x$ mole ratio in the flue gas of at least 3-5 is required for maximum performance. This can easily be shown by observing reactions 2-24 and 2-26 reprinted below.

$$SO_2(aq) + 2CH_3COONa(aq) + H_2O \rightarrow Na_2SO_3(aq) + 2CH_3COOH(aq)$$

$$2NO(aq) + 5Na_2SO_3(aq) + 4CH_3COOH(aq) \rightarrow 2NH(SO_3Na)_2(aq) + Na_2SO_4(aq) + 4CH_3COONa(aq) + H_2O$$

One can see that 1 mole of $SO_2$ absorbed in solution reacts to form 1 mole of sodium sulfite ($Na_2SO_3$). Then, 5 moles of sodium sulfite are required to reduce 2 moles of NO. So, the minimum stoichiometric $SO_2:NO_x$ mole ratio required is $\frac{5}{2}$ or 2.5. Also, some of the sodium sulfite is oxidized to sodium sulfate by oxygen present in the flue gas according to:

$$Na_2SO_3(aq) + \tfrac{1}{2}O_2(aq) \rightarrow Na_2SO_4(aq) \qquad (2\text{-}33)$$

and is not available for $NO_x$ reduction. Low-sulfur coals would require $SO_2$ to be added to the flue gas for these processes to perform; therefore, they should be considered applicable to high sulfur coals only.

Absorption-reduction processes require large absorbers with high liquid rates due to relative insolubility of NO, even when the absorption catalyst is used. Also, the regeneration of the absorption catalyst and the flue gas reheat for plume buoyancy are energy intensive. Some corrosion-resistant material is necessary due to the corrosive nature of the absorbing solution. However, absorption-reduction appears to be the most promising of the "wet" $NO_x/SO_2$ removal processes. This is due primarily to its not utilizing oxidants which require much corrosion-resistant material and, more importantly, create serious secondary pollution problems. Also, the primary by-products of absorption-reduction processes, gypsum, can be used as landfill material or in building materials. For all the above reasons, absorption-reduction processes appear, at this preliminary stage, competitive with other wet $NO_x/SO_2$ removal processes.

### 2.2.9 Oxidation-Absorption-Reduction

#### 2.2.9.1 System Description--

Oxidation-absorption-reduction processes simultaneously remove $NO_x$ and $SO_2$ from flue gas by oxidizing relatively insoluble NO to relatively soluble $NO_2$ and then absorbing both $NO_2$ and $SO_2$ into a scrubbing solution. The processes are based on the use of gas-phase oxidants, either ozone ($O_3$) or chlorine dioxide ($ClO_2$), to selectively oxidize NO to $NO_2$. Both perforated-plate and packed bed absorption columns are utilized by various process vendors.

Most of the oxidation-absorption-reduction processes are similar in that they consist of five major sections:

- prescrubbing
- gas-phase oxidation
- $NO_2/SO_2$ absorption
- reduction of absorbed $NO_x$ and oxidation of $SO_3^=$
- wastewater treatment

The areas where processes differ are gas-phase oxidation - $O_3$ or $ClO_2$; absorption solutions - limestone slurry ($CaCO_3$), $H_2SO_4$, or NaOH; and the amount and type of waste treatment required. Thermal decomposition, biological denitrification, or wastewater evaporation wastewater treatment systems can be used. Because of these differences, only one of the oxidation-absorption-reduction processes, that of Mitsubishi Heavy Industries, is examined here in detail.

A block flow diagram of the MHI oxidation-absorption-reduction process is shown in Figure 2.2.9-1.

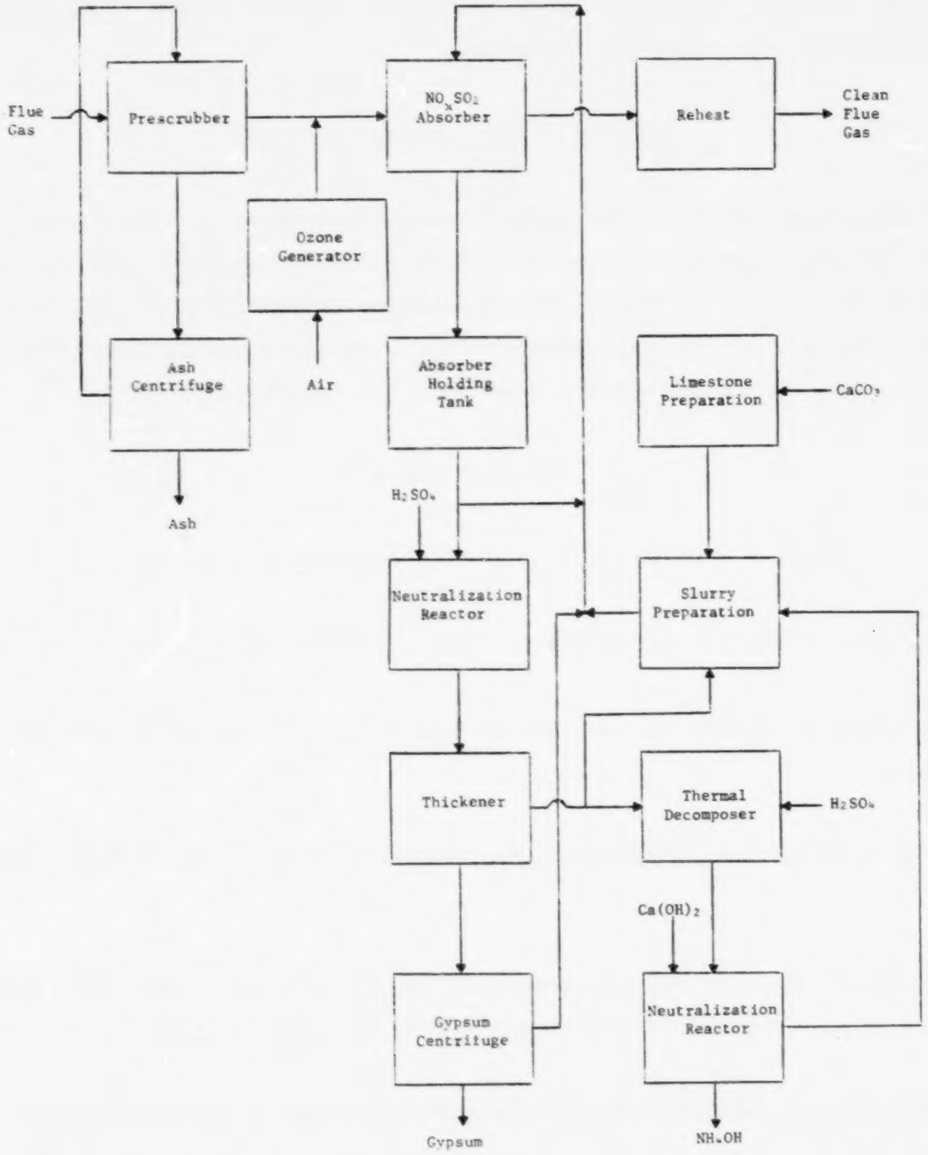

Flue Gas
Prescrubber
$NO_x SO_2$ Absorber
Reheat
Clean Flue Gas
Ozone Generator
Air
Ash Centrifuge
Ash
Absorber Holding Tank
Limestone Preparation
$CaCO_3$
$H_2SO_4$
Neutralization Reactor
Slurry Preparation
Thickener
Thermal Decomposer
$H_2SO_4$
$Ca(OH)_2$
Gypsum Centrifuge
Neutralization Reactor
Gypsum
$NH_4OH$

Figure 2.2.9-1. Process flow diagram for MHI oxidation-absorption-reduction process.[81]

Flue gas is taken from the boiler after the air preheater and passed through a prescrubber to cool the gas and remove particulates and chlorides. The flue gas then enters a duct where it is injected with ozone (about 1 percent by weight in air)[82] such that the $O_3$:NO ratio is 1:1. Ozone selectively oxidizes NO by the following reatcion.[83]

$$NO(g) + O_3(g) \rightarrow NO_2(g) + O_2(g) \tag{2-34}$$

After injection, the flue gas passes countercurrent to a lime/limestone slurry in a grid-packed absorption column. A water-soluble catalyst is added to the slurry to enhance $NO_2$ absorption (even though $NO_2$ is more soluble than NO, it is still less soluble than $SO_2$). $SO_2$ is absorbed quickly at the bottom of the column and undergoes the following reactions.[15]

$$SO_2(g) \rightarrow SO_2(aq) \tag{2-23}$$

$$SO_2(aq) + CaCO_3(s) + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O(s) + CO_2(g) \tag{2-35}$$

$$SO_2(aq) + CaSO_3(aq) + H_2O \rightarrow Ca(HSO_3)_2(aq) \tag{2-36}$$

$NO_2$ is absorbed gradually over the length of the column and reacts as follows.[16]

$$2NO_2(g) + Ca(OH)_2(s) + CaSO_3 \cdot \tfrac{1}{2}H_2O(s) + \tfrac{1}{2}H_2O \rightarrow Ca(NO_2)_2(aq) + CaSO_4\ 2H_2O(s) \tag{2-37}$$

Once both the $NO_2$ and $SO_2$ are absorbed, the nitrite ion formed by reaction 2-37 is reduced by the bisulfate ion formed by reaction 2-36.[84]

$$Ca(NO_2)_2(aq) + 3Ca(HSO_3)_2(aq) \rightarrow 2Ca[NOH(SO_3)_2](aq) + 2CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \downarrow + H_2O \tag{2-38}$$

These hydroxylamine $[NOH(SO_3)_2^{=}]$ compounds are reduced further by the sulfite ion[85]

$$Ca[NOH(SO_3)_2](aq) + CaSO_3 \cdot \tfrac{1}{2}H_2O(s) + \frac{3}{2}\,H_2O \rightarrow Ca[NH(SO_3)_2](aq) + CaSO_4 \cdot 2H_2O(s)\downarrow \qquad (2\text{-}39)$$

Upon leaving the top of the absorber, the clean flue gas is reheated for plume buoyancy and sent to the stack. The slurry solution drops to a holding tank from which most of the solution is returned to the top of the absorber. A small stream passes to a neutralization reactor where sulfuric acid is added to convert the sulfite solid to soluble bisulfite and solid gypsum:[86]

$$2CaSO_3 \cdot \tfrac{1}{2}H_2O(s) + H_2SO_4(aq) + H_2O \rightarrow CaSO_4 \cdot 2H_2O(s)\downarrow + Ca(HSO_3)_2(aq) \qquad (2\text{-}40)$$

This stream passes to a thickener from which the bottoms are sent to a centrifuge to separate the solid gypsum by-product from the liquor which is returned to the absorber. The overflow from the thickener is primarily recycled to the limestone slurry preparation tank. The remainder is sent to a thermal decomposer where sulfuric acid is added to hydrolyze the N-S compounds.[18]

$$2Ca[NH(SO_3)_2](aq) + 2H_2O \overset{H^+}{\rightarrow} Ca(NH_2SO_3)_2(aq) + Ca(HSO_4)_2(aq) \qquad (2\text{-}41)$$

$$Ca(NH_2SO_3)_2(aq) + Ca(HSO_4)_2(aq) + 6H_2O \overset{H+}{\rightarrow} 2NH_4HSO_4(aq) + 2CaSO_4 \cdot 2H_2O(s)\downarrow \qquad (2\text{-}42)$$

The ammonium bisulfate solution is pumped to another neutralization reactor where lime is added.[87]

$$NH_4HSO_4(aq) + Ca(OH)_2 + H_2O \rightarrow CaSO_4 \cdot 2H_2O(s)\downarrow + NH_4OH(aq) \qquad (2\text{-}43)$$

MHI has three possible methods of removing this ammonium hydroxide:

- decompose by increasing pH
- decompose thermally
- strip out with makeup $H_2O$

The remaining gypsum slurry is pumped to the limestone slurry preparation tank.

The fundamental design equation used for gas absorption column design is[32]

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \qquad (2\text{-}11)$$

where $y$ = bulk $NO_x$ concentration (mole fraction) of gas phase at any given point in column

$y-y^*$ = overall driving force for absorption ($y^*$ being the $NO_x$ concentration of a gas in equilibrium with a given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/($ft^2$)(hr)(mole fraction)

$a$ = area of gas-liquid interface per unit packed volume, $ft^2/ft^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/($ft^2$)(hr)

$Z$ = length of packed section of column, ft

In a column containing a given plate or packing configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity obviously, must be lower than the flooding velocity. How much lower is a design decision. Often, it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[33] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter consideration is why oxidation-absorption-reduction processes oxidize NO to more soluble $NO_2$ prior to the absorber and why some processes add water soluble catalysts to the scrubbing solution to aid $NO_2$ absorption. The oxidation step enables these processes to use shorter absorbers with lower liquid rates than either the absorption-oxidation or absorption-reduction processes. Table 2.2.9-1 presents the effects of boiler/flue gas variables on the design of oxidation-absorption-reduction systems. Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load change. The column size and the liquid, oxidant, and catalyst flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

Typical ranges for several operating parameters for this type of process are shown in Table 2.2.9-2. Reagent concentrations were not available. Economics for the various processes cover a wide range presumably due to different techniques for oxidant generation and treatment of the scrubbing solution. Costs are reported to range from $84-134/kW for utility applications' capital expense and 6.7-9 mills/kWh for operating expense.[91]

Presently, some of the oxidation-absorption-reduction processes have reached the prototype stage of development. Table 2.2.9-3 presents a list of oxidation-absorption-reduction process vendors and the status of development of their projects. One can see from the table that no coal-fired flue gas tests have been made as of yet.

TABLE 2.2.9-1. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Depends on individual process: if $NO_2^-$ is completely reduced to $N_2$ or $NH_3$ by $SO_3^=$ (as does MHI), then at least the stoichiometric $SO_2:NO_x$ mole ratio of 3:1 is required[88] [see equation (9-6)]; if $NO_2^-$ is not reduced completely, then a different ratio will be necessary |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased gas-phase oxidant flow rate; increased liquid-phase catalyst concentration |

TABLE 2.2.9-2. TYPICAL RANGES OF OPERATING VARIABLES FOR OXIDATION-ABSORPTION-REDUCTION PROCESSES[89,90]

| Variable | | Range |
|---|---|---|
| Liquid/Gas Ratio, $1/Nm^3$ | | 2-12 |
| Oxidant/NO Mole Ratio | $O_3$ systems | 0.6-1.0 |
| | $ClO_2$ systems | 0.55 |
| $SO_2/NO_x$ Mole Ratio | | 2.5-5.0 |
| Superficial Gas Velocity, m/s | | 3-5 |
| Pressure Drop, $mmH_2O$ | | 200-500 |

TABLE 2.2.9-3. PROCESS VENDORS OF OXIDATION-ABSORPTION-REDUCTION PROCESSES[92,93]

| Vendor | Status of Development |
|---|---|
| Chiyoda | 1975: 1000 $Nm^3$/hr flue gas from heavy oil-fired boiler |
| Ishikawajima-Harima Heavy Industries | 1975: 5000 $Nm^3$/hr flue gas from oil-fired boiler (3000 hours continuous) |
| Mitsubishi Heavy Industries | 1975: 2000 $Nm^3$/hr flue gas from heavy oil-fired boiler (700 hours continuous) |
| Osaka Soda | 1976: 60,000 $Nm^3$/hr flue gas from oil-fired boiler |
| Shirogane | 1974: 48,000 $Nm^3$/hr flue gas from oil-fired boiler |
| Sumitomo Metal-Fujikasui: Calcium Process | 1976: 25,000 $Nm^3$/hr flue gas from sintering furnace |
| Sumitomo Metal-Fujikasui: Sodium Process | 1973: 62,000 $Nm^3$/hr flue gas from heavy oil-fired boiler (5 others) |

2.2.9.2 System Performance--

No coal-fired testing has been performed. Results of oil-fired tests show up to 90 percent $NO_x$ reduction and >95 percent $SO_2$ reduction.

The primary disadvantage of these processes is the utilization of costly gas-phase oxidants which create secondary wastewater pollution problems. Both ozone and chlorine dioxide are highly unstable so they cannot be stored and must be generated onsite. $O_3$, the more expensive of the two, is generated by a high energy corona discharge in air. This instantaneous process requires significantly large amounts of electricity. $ClO_2$ is generated by a slower chemical reaction (requires about 20 minutes to respond to a change in demand) which could make it less responsive to boiler load changes. The use of $ClO_2$ introduces an additional secondary pollutant, chlorides, besides the nitrite salt problem. Significant amounts of corrosion-resistant material are required for oxidation-absorption-reduction processes, regardless of which oxidant is utilized. Some of the processes would not be applicable to low sulfur coals as they require large amounts of $SO_2$ to obtain $NO_2(aq)$ or $NO_2^-$ reduction.

## 2.2.10 Oxidation-Absorption

### 2.2.10.1 System Description--

As a group, oxidation-absorption processes include those oxidation processes which do not qualify for the oxidation-absorption-reduction category. Basically, there are two types of oxidation-absorption processes. One is a simplified version of the oxidation-absorption-reduction process and uses an excess of ozone to selectively oxidize $NO_x$ to $N_2O_5$ which is absorbed into aqueous solution and concentrated to form a 60 percent nitric acid ($HNO_3$) by-product. There is no reduction of $NO_x(NO_2^-)$ by the absorption of $SO_2$(as $SO_3^=$) and no wastewater treatment facility. The other type of oxidation-absorption process is based on equimolar $NO$-$NO_2$ absorption: absorbing $N_2O_3$ which is formed by the gas-phase reaction of $NO$ and $NO_2$.

A flow diagram of the Kawasaki Heavy Industries oxidation-absorption process is shown in Figure 2.2.10-1. Flue gas is taken from the boiler after the air preheater. It passes countercurrent to a magnesium hydroxide [$Mg(OH)_2$] slurry in the first section of the absorber. There, $SO_2$ is absorbed and undergoes the following reactions.[95]

$$SO_2(g) \rightarrow SO_2(aq) \qquad (2\text{-}23)$$

$$Mg(OH)_2(s) + SO_2(aq) + 5H_2O \rightarrow MgSO_3 \cdot 6H_2O(s)\downarrow \qquad (2\text{-}44)$$

The $SO_2$-free flue gas passes to the first denitrification section of the absorber while the liquid effluent drops to a holding tank. A recycle $NO_2$ stream is added to the flue gas to bring the $NO:NO_2$ mole ratio to 1:1. The

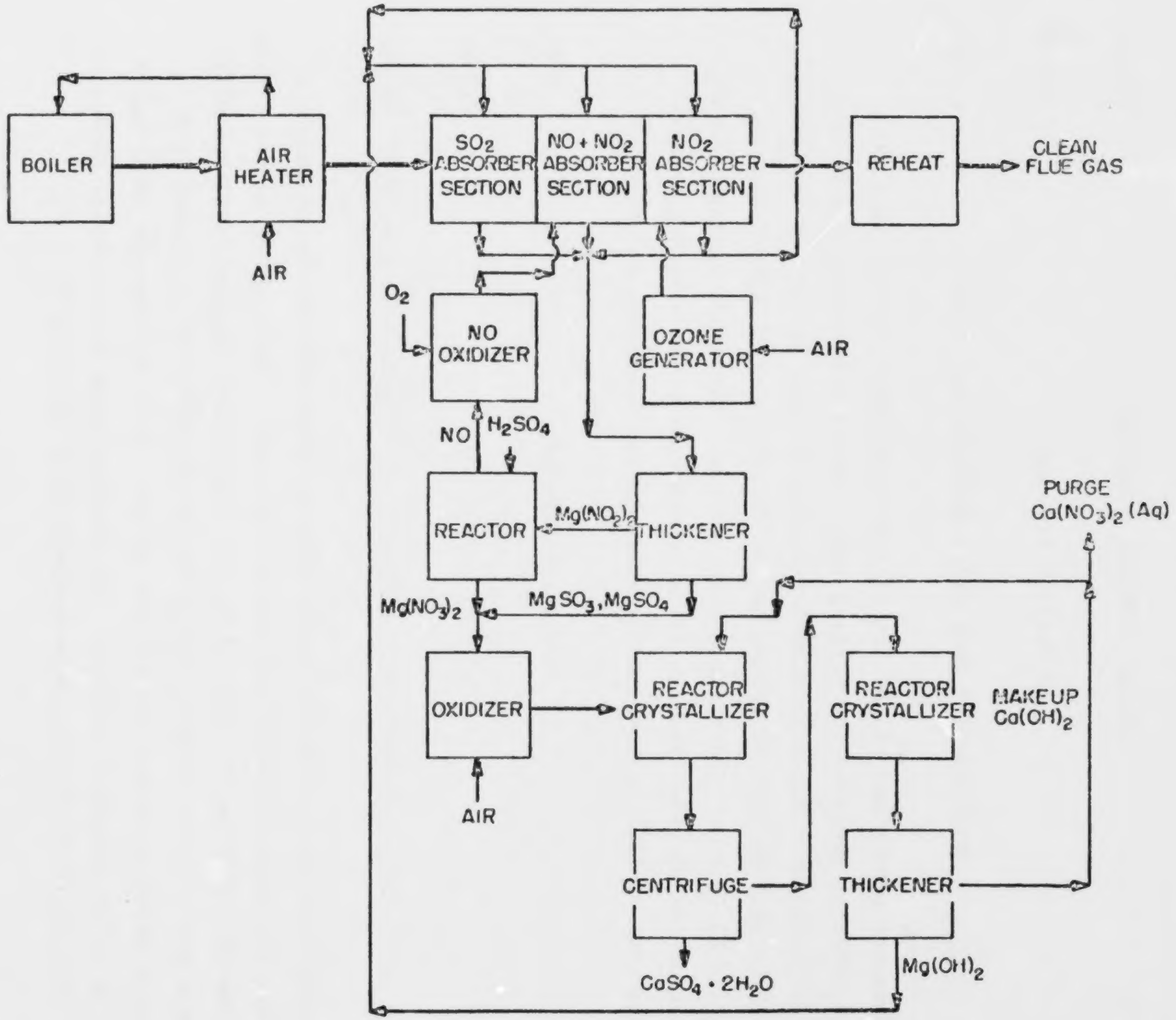

BOILER
AIR HEATER
AIR
$SO_2$ ABSORBER SECTION
$NO + NO_2$ ABSORBER SECTION
$NO_2$ ABSORBER SECTION
REHEAT
CLEAN FLUE GAS
$O_2$
NO OXIDIZER
OZONE GENERATOR
AIR
NO
$H_2SO_4$
REACTOR
$Mg(NO_2)_2$
THICKENER
PURGE
$Ca(NO_3)_2$ (Aq)
$Mg(NO_3)_2$
$MgSO_3, MgSO_4$
OXIDIZER
REACTOR CRYSTALLIZER
REACTOR CRYSTALLIZER
MAKEUP
$Ca(OH)_2$
AIR
CENTRIFUGE
THICKENER
$CaSO_4 \cdot 2H_2O$
$Mg(OH)_2$

Figure 2.2.10-1. Flow diagram of Kawasaki Heavy Industries process.[94]

resulting mixture then passes countercurrent to a $Mg(OH)_2$ slurry. Equimolar amounts of NO and $NO_2$ react and are absorbed in the following manner.[96]

$$NO(g) + NO_2(g) \rightarrow N_2O_3(g) \tag{2-45}$$

$$N_2O_3(g) \rightarrow N_2O_3(aq) \tag{2-46}$$

$$Mg(OH)_2(aq) + N_2O_3(aq) \rightarrow Mg(NO_2)_2(aq) + H_2O \tag{2-47}$$

The flue gas passes out of the top of this absorption section while the liquid effluent drops to the holding tank. Because the rate of reaction 2-45 decreases with $NO_x$ concentration (below 200 ppm it becomes negligible), it is necessary to further reduce $NO_x$ by injecting ozone to oxidize the remaining NO to $NO_2$. The mixture then passes to the final denitrification section of the absorber and is passed countercurrent to a $Mg(OH)_2$ slurry. This section of the absorber is described by the following reactions.[97]

$$2NO_2(g) \rightarrow N_2O_4(g) \tag{2-8}$$

$$N_2O_4(g) \rightarrow N_2O_4(aq) \tag{2-9}$$

$$2N_2O_4(aq) + 2Mg(OH)_2(s) \rightarrow Mg(NO_3)_2(aq) + Mg(NO_2)_2(aq) + 2H_2O \tag{2-48}$$

The clean flue gas leaves the top of this absorber section, is passed to a reheater for plume buoyancy and sent to the stack. Part of the liquid effluent from this section is recycled to the tops of the absorber sections while the rest drops to the holding tank. The slurry solution is pumped to a thickener which separates the soluble nitrite ($NO_2^-$) and nitrate ($NO_3^-$) salts from the solid magnesium sulfite. The overflow from the thickener passes to a $NO_2^-$ decomposition reactor where sulfuric acid is added.[98]

$$3Mg(NO_2)_2(aq) + 2H_2SO_4(aq) \rightarrow 2MgSO_4(aq) + Mg(NO_3)_2(aq) + 4NO(g)\uparrow + 2H_2O \tag{2-49}$$

The NO off-gas passes through an oxidizer where it is oxidized by air to $NO_2$ and sent to the first denitrification section of the absorber. The effluent from the decomposition reactor is mixed with the thickener bottoms and pumped to a second oxidizer.[99]

$$MgSO_3 \cdot 6H_2O(s) + \tfrac{1}{2}O_2(g) \rightarrow MgSO_4(aq) + 6H_2O \qquad (2\text{-}50)$$

The magnesium sulfate formed in the oxidizer is treated with calcium nitrate in a gypsum production reactor.[25]

$$Ca(NO_3)_2(aq) + MgSO_4(aq) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \downarrow + Mg(NO_3)_2(aq) \qquad (2\text{-}51)$$

The products of this reaction are sent to a centrifuge to remove the solid gypsum by-product. The liquid from the centrifuge goes to another decomposition reactor where makeup lime slurry is added.[100]

$$Mg(NO_3)_2(aq) + Ca(OH)_2(s) \rightarrow Ca(NO_3)_2(aq) + Mg(OH)_2(s) \qquad (2\text{-}52)$$

The magnesium hydroxide product is separated in a thickener and recycled to the absorbers. The thickener overflow stream is split and part is recycled to the gypsum production reactor and the rest is concentrated to form low-grade liquid fertilizer by-product, $Ca(NO_3)_2$.

Since the processes in this category are all very different, especially with respect to chemistry, generalization of typical ranges of operating variables is not meaningful and, therefore, not presented. No published economics for these processes were found.

Presently, the equimolar absorption-type oxidation-absorption processes are still in the pilot-unit stage of development. Table 2.2.10-1 presents a list of all oxidation-absorption process vendors and their project's status of development.

TABLE 2.2.10-1. PROCESS VENDORS OF OXIDATION-ABSORPTION PROCESSES[100]

| Vendor | Status of Development |
|---|---|
| Kawasaki Heavy Industries | 1975: 5000 $Nm^3$/hr flue gas from coal-fired boiler |
| Tokyo Electric-MHI ($NO_x$ only) | 1974: 100,000 $Nm^3$/hr flue gas from natural gas-fired boiler |
| Ube Industries | No information available |

2.2.10.2 System Performance--

Only one coal-fired test has been performed. No information has been published on any of the tests conducted.

The production of nitrate salts poses a potential secondary pollution problem. The plan for reclaiming and concentrating the nitrates as $Ca(NO_3)_2(aq)$ for liquid fertilizer is questionable as the by-product is of low quality and may not be easily marketable in the U.S. Also, the gypsum by-product would be contaminated with various nitrate and sulfite salts, and therefore, would probably be useful only as landfill material. Much corrosion-resistant material is necessary due to the utilization of ozone and circulating magnesium slurries. The three absorber sections, with their respective operating conditions, and ozone generation present complex process control problems. The process steps of several absorber sections in series (large fan requirements), ozone generation (corona discharge), flue gas reheat (inline heater), and by-product and wastewater treatment are all energy intensive and present technical and economic disadvantages when compared to other simpler FGT processes.

## 2.3 CONTROLS FOR OIL-FIRED BOILERS

### 2.3.1 Selective Catalytic Reduction-Fixed Packed Bed Reactors

#### 2.3.1.1 System Description--

Fixed packed bed systems are applicable only to flue gas streams containing less than 20 mg/Nm$^3$ of particulates. As such, they are applicable to distillate oil-fired boilers (19 mg/Nm$^3$) but not to residual oil-fired boilers (330 mg/Nm$^3$).

The primary feature of these systems is the reactor which contains the catalyst. As the name implies, the granular catalyst is randomly packed in a stationary bed. An example of a typical fixed bed reactor is shown in Figure 2.3.1-1. The important features of the reactor are:

- the catalyst
- the catalyst support
- the gas distributor

The catalyst can be either spherical or cylindrical in shape. Spherical granules typically range in size from 4-10 mm in diameter.[103] The composition varies from process to process and most formulations are proprietary. All of the catalysts considered here for use in treating flue gas containing $SO_2$ and $SO_3$ are resistant to poisoning by these compounds. Long term tests of these catalysts in the presence of $SO_x$ have shown very little or no decrease in activity or selectivity. The catalyst is supported either by inert packing (as shown in Figure 2.3.1-1) or by a perforated support plate (Figure 2.3.1-2).

The catalyst supports hold the catalyst fixed in place in order to prevent both mobilization of the particles by the gas stream and catalyst rearrangement which would allow channelling of the flue gas. The gas distributor can be a perforated plate or similar device which spreads the gas flow across the entire cross-section of the catalyst bed.

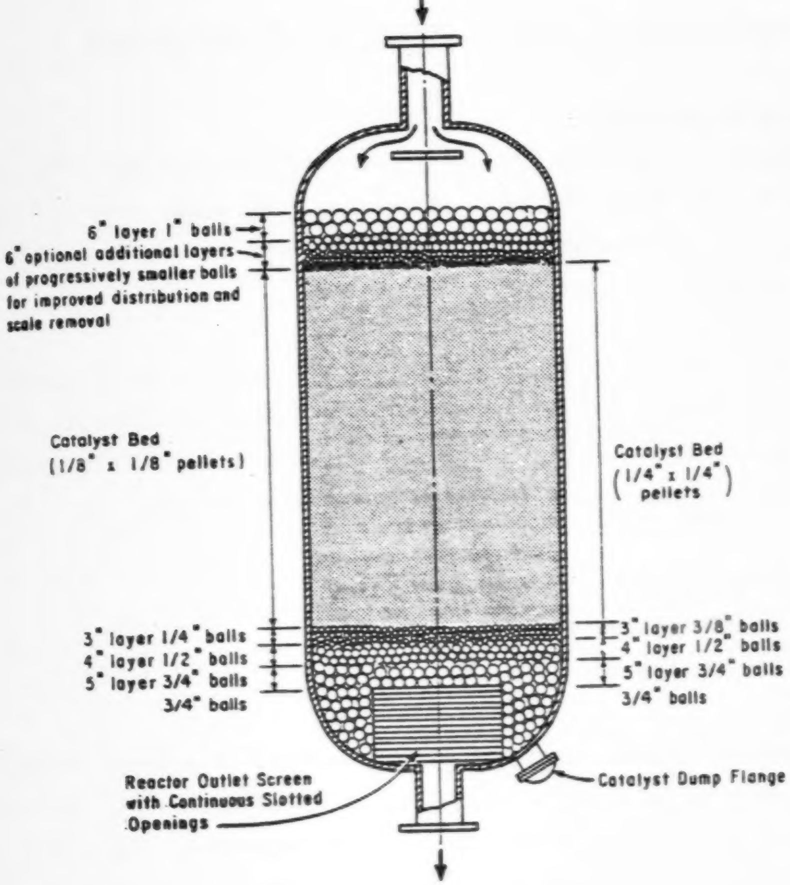

6" layer 1" balls
6" optional additional layers of progressively smaller balls for improved distribution and scale removal
Catalyst Bed (1/8" x 1/8" pellets)
Catalyst Bed (1/4" x 1/4" pellets)
3" layer 1/4" balls
4" layer 1/2" balls
5" layer 3/4" balls
3/4" balls
3" layer 3/8" balls
4" layer 1/2" balls
5" layer 3/4" balls
3/4" balls
Reactor Outlet Screen with Continuous Slotted Openings
Catalyst Dump Flange

Figure 2.3.1-1. Example of typical fixed packed bed reactor.[101]

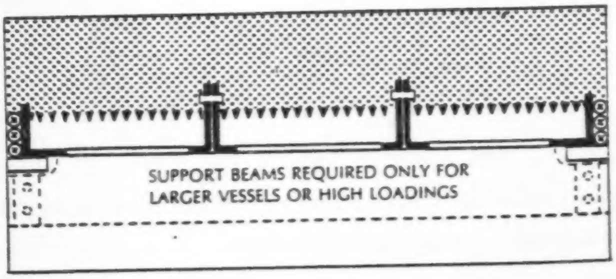

SUPPORT BEAMS REQUIRED ONLY FOR LARGER VESSELS OR HIGH LOADINGS

Figure 2.3.1-2. Example of catalyst support plate.[102]

A typical fixed bed SCR process layout is presented in Figure 2.3.1-3. Several arrangements are possible, however, for application to new boilers this arrangement is the most desirable.[8]

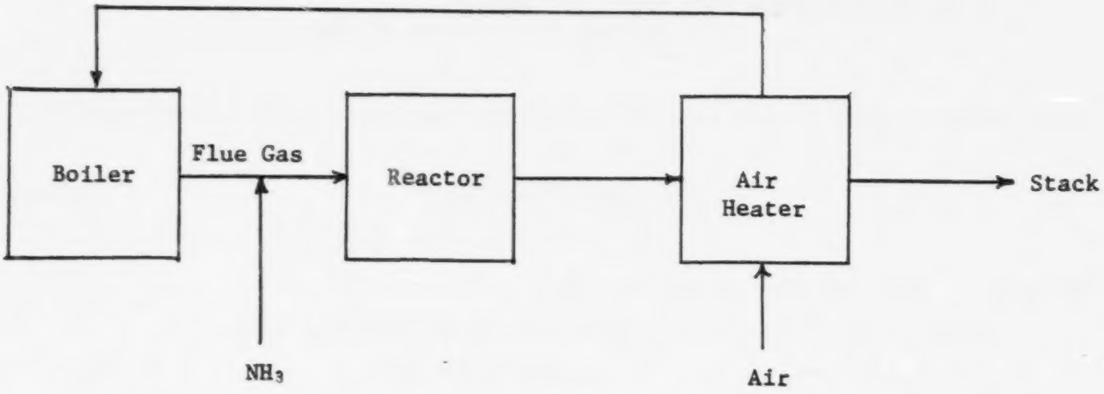


Figure 2.3.1-3. Process layout for fixed bed SCR process.

The principle of operation of these systems involves a gas phase reaction between ammonia ($NH_3$) and $NO_x$ ($NO$ and $NO_2$). These reactions are presented most accurately by[12]

$$4NO + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \qquad (2\text{-}1)$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \qquad (2\text{-}2)$$

The first reaction predominates since flue gas $NO_x$ is typically 90-95 percent NO. As shown, the $NO_x$ is reduced to molecular nitrogen ($N_2$) which exits with the flue gas stream.

The primary design equation used with these processes is the standard equation for reactor design,[13]

$$\frac{V}{F} = \int_o^x \frac{dx}{r} \qquad (2\text{-}3)$$

where V is the catalyst volume

F is the mass (or molar) flow rate

x is the conversion of $NO_x$ to $N_2$

r is the reaction rate $\frac{\text{mass (or moles)}}{\text{volume of catalyst x time}}$

The reaction rate, r, for each NO reduction reaction can be represented by

$$r = k[NH_3]^a[NO]^b[O_2]^c \tag{2-4}$$

where k is the reaction rate constant

$[NH_3]$, $[NO]$, $[O_2]$ are the reactant concentrations, and

a, b, c are empirically determined exponents

The catalyst volume can also be determined if the space velocity is known for the catalyst and removal level of interest. The space velocity is defined as the gas flow rate divided by the catalyst volume. The reaction rate is different for each catalyst formulation and therefore, values for k, a, b, and c must be determined for the particular catalyst to be used before any design can be performed. The reaction rate constant is usually described by the Arrhenius equation.

$$k = Ae^{-\frac{E}{RT}} \tag{2-5}$$

where A is the frequency factor

E is the activation energy

R is the universal gas constant, and

T is the temperature

Values for k, a, b and c for two catalyst formulations are shown in Table 2.3.1-1. Values for other catalyst formulations will be different. The most important design and operating variables are similar to those for moving bed systems using granular catalysts. These are listed, along with typical ranges, in Table 2.3.1-2.

Other variables that affect the process are

- flue gas flow rate
- $NO_x$ control level
- $NO_x$ concentration
- boiler load variation

The flue gas flow rate and control level determine the catalyst volume (hence reactor size) necessary. Increases in either also increases the reactor size. The $NO_x$ concentration is a function of fuel type used in the standard boilers. Higher concentrations require larger $NH_3$ storage and vaporization equipment; reactor size is not affected. Boiler load can affect several things including flue gas temperature, flow rate and $NO_x$ concentration. It is necessary to maintain reactions temperatures of 350 to 400°C and temperature control equipment may be necessary if the boiler experiences large load variations. Where these variations are present, some equipment overdesign may be warranted to insure a constant control level. These variables are discussed in more detail in the section on moving bed SCR systems for coal-fired boilers, Section 2.3.2. Costs of fixed packed bed systems range from $16-49/kW (capital) and 1.2-1.8 mills/kWh (operating). These costs are based on utility applications as well as a variety of processes and operating conditions.

There are vendors of fixed packed bed SCR systems and all are Japanese. Vendors are listed in Table 2.3.1-3 and the scale of development is also noted. Fixed packed systems have been applied to industrial but not utility boilers in Japan. Existing and planned installations are shown in Table 2.3.1-4. Currently, there are no installations in the U.S.

## TABLE 2.3.1-1. REACTION RATE DATA FOR TWO CATALYST FORMULATIONS[11]

Catalyst: $V_2O_5$ on $Al_2O_3$

$$k = 2.05 \times 10^3 e^{-\frac{9650}{RT}}$$

$a = 0.30$

$b = 0.22$

$c = 0.05$

Catalyst: Fe-Cr on $Al_2O_3$

$$k = 3.25 \times 10^3 e^{-\frac{10,860}{RT}}$$

$a = 0.45$

$b = 0.10$

$c = 0.15$

## TABLE 2.3.1-2. DESIGN AND OPERATING VARIABLES FOR FIXED PACKED BED SYSTEMS[14]

| Variable | Typical Range |
|---|---|
| Gas Velocity, m/s | 1 - 1.5 |
| Bed Depth, m | 0.2 - 0.6 |
| Space Velocity, $hr^{-1}$ | 6,000 - 10,000 |
| Pressure Drop, $mmH_2O$ | 40 - 80 |
| Temperature, °C | 350 - 400 |

TABLE 2.3.1-3. VENDORS OF SCR FIXED BED SYSTEMS FOR OIL-FIRED APPLICATIONS[21]

| Vendor | Notes |
|---|---|
| Sumitomo Chemical | Tested on commercial scale equipment |
| Hitachi Zosen | Tested on commercial scale equipment |
| Hitachi, Ltd. | Tested on commercial scale equipment |
| Mitsubishi Heavy Industries | Tested on commercial scale equipment |
| Ishikawjima-Harima Heavy Industries | Tested on commercial scale equipment |
| Mitsui Toatsu Chemical | Has not been applied to boilers |
| Kawasaki Heavy Industries | Tested on pilot scale equipment |
| Mitsubishi Kakoki Kaisha | Tested on commercial scale equipment |

#### 2.3.1.2 System Performance--

Typical performance data for packed fixed bed SCR systems are shown in Figures 2.3.1-4 and 2.3.1-5 and Tables 2.3.1-5 through 2.3.1-7. These data indicate that $NO_x$ removals up to 90 percent are achievable with these systems. This allows them to be considered for all control levels of interest in this study.

There are some potential problems downstream of the SCR systems (fixed packed bed, moving, and parallel flow) due to the presence of the unreacted ammonia in the flue gas. Two things can happen: 1) the $NH_3$ can react with $SO_2$ or $SO_3$ to form ammonium bisulfate or ammonium sulfate or 2) the $NH_3$ can enter the downstream equipment unreacted. The bisulfate has been shown to cause air preheater pluggage and this is the subject of ongoing research both at the EPA and the Electric Power Research Institute (EPRI). Both the bisulfate and sulfate exist as a particulate, but may be difficult to collect if the particles are submicron in size. Unreacted $NH_3$ is not likely to present any operational problems. A recent study has shown that if an ESP exists downstream, then most of the $NH_3$ will exit with the ash. $NH_3$ can actually improve the performance of an FGD system.[129]

TABLE 2.3.1-4. EXISTING FGT INSTALLATIONS OF SCR FIXED BED SYSTEMS OIL-FIRED INDUSTRIAL BOILERS[21]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Amagasaki | Kansai Paint | Hitachi, Ltd. | Distillate | 16,000 | October 1977 |
| Amagasaki | Nisshin Steel | Hitachi, Ltd. | Resid | 20,000 | August 1977 |
| Amagasaki | Nisshin Steel | Hitachi, Ltd. | Resid | 19,000 | July 1977 |
| Sakai | Nisshin Steel | Hitachi, Ltd. | Distillate | 30,000 | December 1978 |
| Hokkaichi | Shindaikyowa P.C. | Hitachi Zosen | Resid | 440,000 | November 1975 |
| Sodegaura | Sumitomo Chemical | Sumitomo Chemical Eng. | Resid | 30,000 | July 1973 |
| Sodegaura | Sumitomo Chemical | Sumitomo Chemical Eng. | Resid | 240,000 | March 1976 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 62,000 | October 1977 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 23,000 | December 1977 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 23,000 | June 1978 |
| Sorami | Toho Gas | Sumitomo Chemical Eng. | Distillate | 19,000 | July 1978 |
| Kawasaki | Nippon Yakin | Mitsubishi Kakoki | Resid | 14,000 | July 1978 |
| Kawasaki | Toho Gas | Mitsubishi Kakoki | Distillate | 30,000 | October 1977 |
| Chita | Toho Gas | Mitsubishi Kakoki | Distillate | 30,000 | October 1977 |

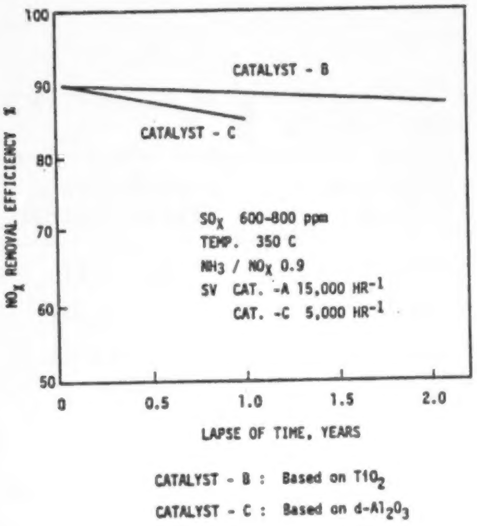


Figure 2.3.1-4. Performance of experimental catalyst of Sumitomo Chemical.[104]

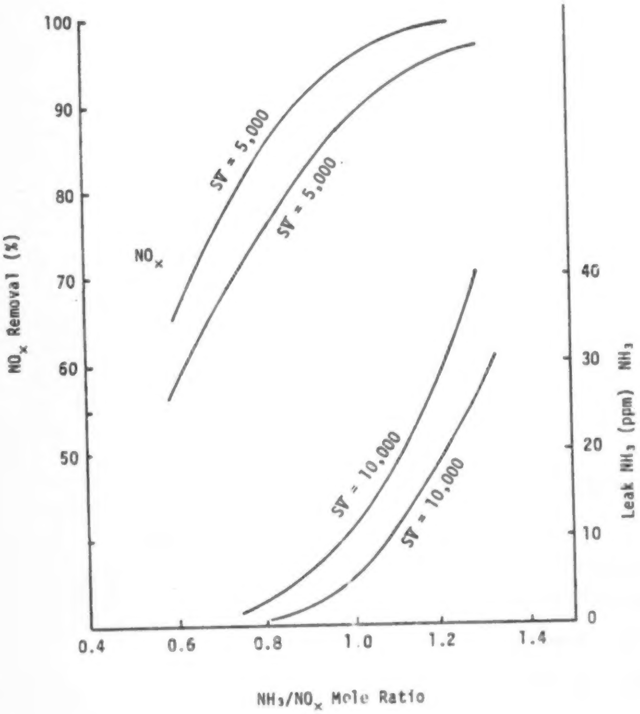


Figure 2.3.1-5. Typical example of operation data (oil-fired boiler, 350-400°C, granular or honeycomb catalyst).[105]

TABLE 2.3.1-5. OPERATION PARAMETERS OF MAJOR PLANTS CONSTRUCTED BY HITACHI ZOSEN[106]

| | Idemitsu Kosan | Shindaikyowa Petrochemical | Kawasaki Steel |
|---|---|---|---|
| Completed | Oct. 1975 | Nov. 1975 | Nov. 1976 |
| Plant site | Chiba | Yokkaichi | Chiba |
| Gas source | FCC-CO Boiler and furnace | Oil-fired Boiler | Iron-ore Sintering machine |
| Capacity ($Nm^3$/hr) | 350,000 | 440,000 | 762,000 |
| Load factor (%) | 50-100 | 50-100 | 70-100 |
| Pretreatment of gas | Heating | EP*, FGD, Heating | EP, FGD<br>WEP†,<br>Heating |
| Reactor inlet | | | |
| $NO_x$ (ppm) | 230 | 150 | 200-300 |
| $SO_x$ (ppm) | 50-80 | 80-130 | 5-20 |
| Dust (mg/$Nm^3$) | 20-50 | 30-100 | 3-10 |
| $O_2$ (%) | 2.3 | 3.2 | 11.2 |
| Reactor type | Fixed bed | Fixed bed | Fixed bed |
| Reaction temp. | 400 | 420 | |
| $NO_x/NH_3$ ratio | 1.0 | 1.0 | 1.0 |
| Catalyst No. | 204 | 304 | 304 |
| SV ($hr^{-1}$) | 5,000 | 10,000 | 4,000 |
| $NO_x$ removal (%) | 93 | 80‡ | 95 |
| Pressure drop by SCR reactor ($mmH_2O$) | 170 | 160 | 50 |
| Catalyst life | 1 year | 1 year | 1 year |

*Electrostatic precipitator

†Wet electrostatic precipitator

‡Including leakage in heat exchanger

TABLE 2.3.1-6. SCR PLANT BY MITSUI ENGINEERING & SHIPBUILDING CO.[107]

| | Mitsui Petro-Chemical Co. |
|---|---|
| Capacity ($Nm^3/hr$) | 200,000 |
| Gas composition | |
| $NO_x$ (ppm) | 190 |
| $SO_x$ (ppm) | None |
| Dust ($mg/Nm^3$) | 20-50 |
| Catalyst and reactor | |
| Catalyst carrier | $Al_2O_3$ |
| Catalyst shape | Granule |
| SV ($hr^{-1}$) | 2,600 |
| Temperature (°C) | 350 |
| $NH_3/NO_x$ mole ratio | 1.0 |
| $NO_x$ removal (%) | Above 90 |
| Total pressure drop ($mmH_2O$) | |
| Leak $NH_3$ (ppm) | |
| Operation start | Oct. 1975 |
| Plant cost ($10^6$ yen) | |
| Denitrification cost (yen/kWhr)* | |

*Including 7 years depreciation.

TABLE 2.3.1-7. OPERATION DATA OF SCR PLANTS FOR DIRTY GAS[108]

| | Pilot | Commercial |
|---|---|---|
| Gas for SCR ($Nm^3/hr$) | 30,000 | 240,000 |
| Fuel | Oil(S=0.7%) | Oil(S=0.7%) |
| Load fluctuation | 60-100% | 60-100% |
| Stack height (m) | 70 | 140 |
| Inlet gas composition | | |
| $O_2$ (%) | 6 | 6 |
| $SO_x$ (ppm) | 400 | 400 |
| $NO_x$ (ppm) | 200 | 200 |
| Particulates after EP ($mg/Nm^3$) | 5-20 | 5-10 |
| FGD unit | None | None |
| SV ($hr^{-1}$) | 5,000 | 5,000 |
| Temperature (°C) | 320 | 320 |
| $NO_x$ removal (%) | Over 90 | Over 90 |
| $NH_3$/NO mole ratio | 1.0 | 1.0 |
| Leak ammonia (ppm) | 10-20 | 10-20 |
| Type of reactor | Fixed bed | Fixed bed |
| Pressure drop ($mmH_2O$) | | |
| Reactor | | 200 |
| Total system | | 500 |
| Plant completed | July 1973 | March 1976 |

## 2.3.2 Selective Catalytic Reduction-Moving Bed Reactor

### 2.3.2.1 System Description--

The primary feature of this and other selective catalytic reduction (SCR) processes is the reactor. The reactor contains the catalyst which allows the reduction reaction to proceed at 350-400°C. In this case the reactor is a moving-bed type in which a portion of the catalyst is either continuously or intermittently removed from service in order to remove the accumulated particulates. Some moving bed reactors are shown in Figure 2.3.2-1. The actual reactor arrangement is highly process specific, however, the principle of operation is the same for all of the processes.

For moderate particulate loadings the bed is moved intermittently and operated as a fixed on bed system most of the time. High particulate loadings require that the bed be moved continuously. Moving bed reactors are reportedly capable of handling up to 2 $g/Nm^3$ of particulates. However, this is more a theoretical than a practical particulate load limit.[9] If possible, this would be sufficient to handle the residual oil-fired boilers (0.33 $g/Nm^3$).

Entrained particulates are generally removed from the catalyst bed by vibrating the bed and screening the catalyst or some other mechanical means. Particulate removal by the bed can be as high as 70 percent.[10]

An example flow diagram for a moving bed SCR process is shown in Figure 2.3.2-2. Flue gas is taken from the boiler between the boiler and the air preheater. An economizer bypass is utilized for temperature control. Ammonia is injected and mixed with the flue gas stream just prior to the reactor. The flue gas passes through the catalyst bed where $NO_x$ is reduced to $N_2$ and is then sent to the stack or further treatment facilities. The catalyst circulates through the reactor and is screened to remove particulates. The particulates are blown to a small baghouse where they are collected.

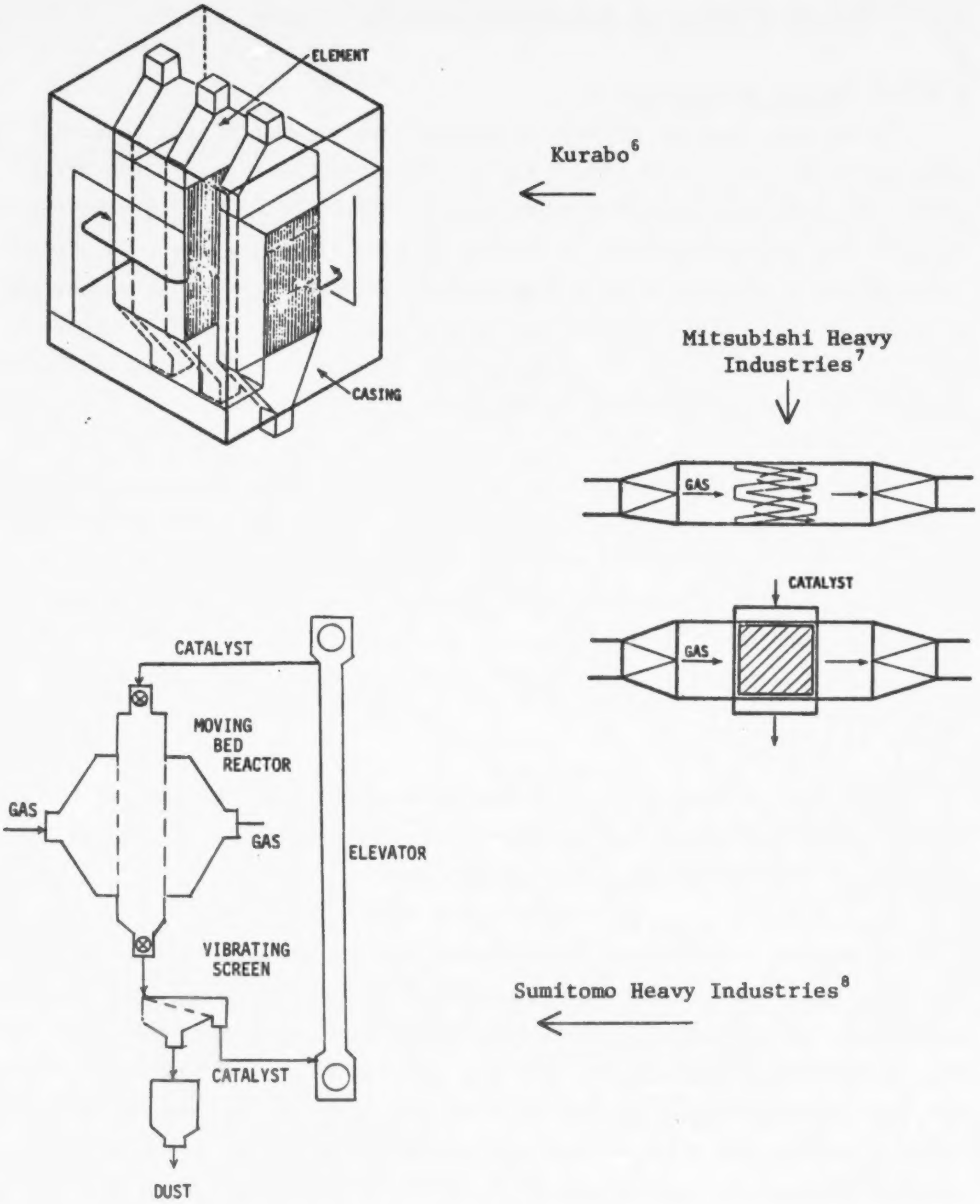

ELEMENT
CASING
Kurabo[6]
Mitsubishi Heavy Industries[7]
GAS
CATALYST
GAS
CATALYST
MOVING BED REACTOR
GAS
GAS
ELEVATOR
VIBRATING SCREEN
CATALYST
DUST
Sumitomo Heavy Industries[8]

Figure 2.3.2-1. Moving bed reactors of three process vendors.

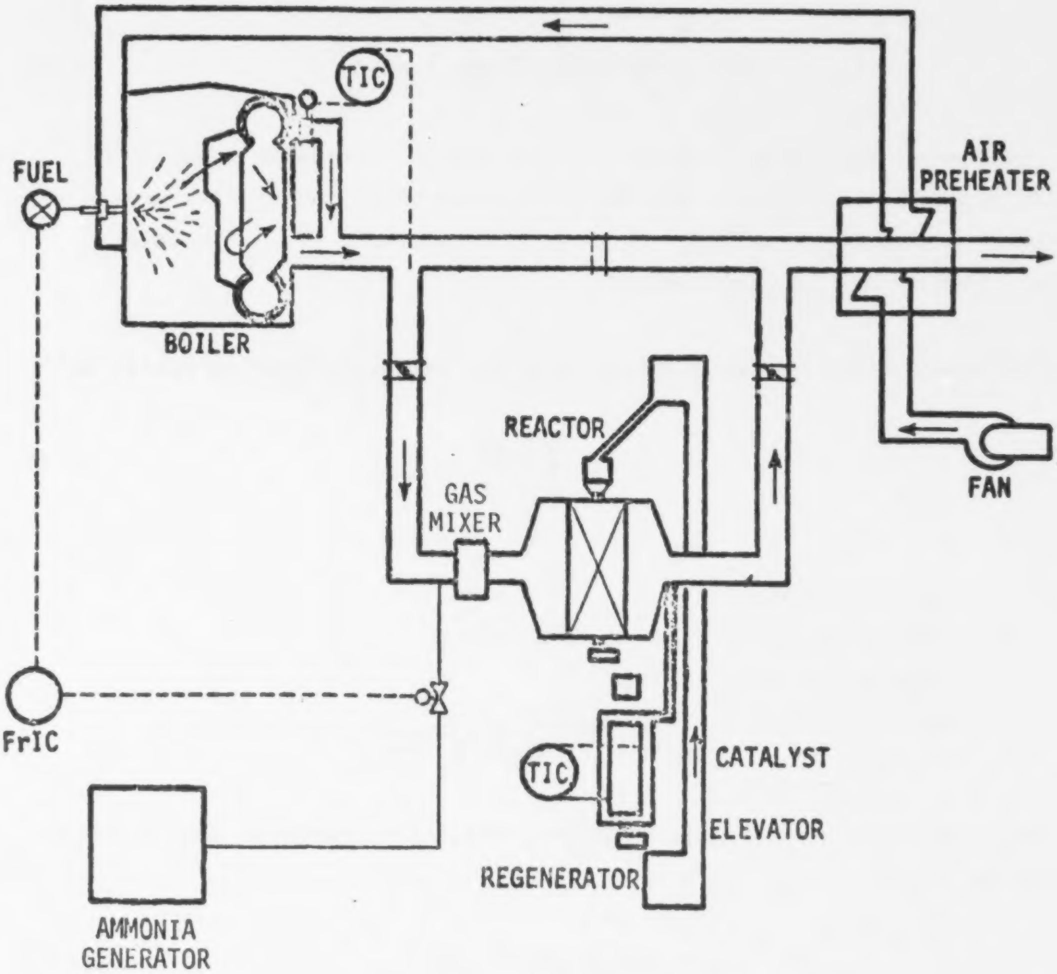

TIC
FUEL
AIR
PREHEATER
BOILER
REACTOR
AIR
FAN
GAS
MIXER
FrIC
TIC
CATALYST
ELEVATOR
REGENERATOR
AMMONIA
GENERATOR

Figure 2.3.2-2. Process flow diagram for moving bed SCR process.[11]

The $NO_x$ reduction reactions are represented most accurately by[12]

$$4NO + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \tag{2-1}$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \tag{2-2}$$

The first reaction predominates since flue gas $NO_x$ typically consists of 90-95 percent NO. As shown, the $NO_x$ is reduced to molecular nitrogen, $N_2$, which exits with the flue gas. $O_2$ is also a reactant, but is in large excess (>3 percent) in the flue gas.

The fundamental design equation used for catalytic reactor design is[13]

$$\frac{V}{F} = \int_{o}^{x} \frac{dx}{r} \tag{2-3}$$

where V = catalyst volume

F = mass flow rate (or molar flow rate)

x = conversion of $NO_x$ to $N_2$

r = reaction rate, $\frac{\text{mass (or moles)}}{\text{volume of catalyst x time}}$

The reaction rate, r, for each of the $NO_x$ reduction reactions can be represented by

$$r = k\ [NH_3]^a\ [NO]^b\ [O_2]^c \tag{2-4}$$

where k = reaction rate constant

$[NH_3]$, $[NO]$, $[O_2]$ = reactant concentrations, and

a, b, c = empirically determined exponents

The reaction rate is different for each catalyst formulation and, therefore, values for k, a, b, and c must be determined for each particular catalyst. The catalyst volume can also be determined if the space velocity is known for the catalyst and removal level of interest. The space velocity is defined as the gas flow rate divided by the catalyst volume.

The most important design and operating variables are presented in Table 2.3.2-1 and typical values for these are also shown. Although some of the data used in developing this table are for utility applications, the values should not be different for industrial applications. There are other variables that are important, but must be determined for each individual case. These are:

- flue gas flow rate
- $NO_x$ control level
- $NO_x$ concentration

TABLE 2.3.2-1. DESIGN AND OPERATING VARIABLES FOR MOVING BED SCR SYSTEMS[14,15]

| Variable | Typical Range |
|---|---|
| Space Velocity | 6000 - 10,000 $hr^{-1}$* |
| $NH_3$:$NO_x$ Mole Ratio | 0.7 - 1.0* |
| Flue Gas Temperature | 350 - 400°C* |
| Pressure Drop | 40 - 80 mm Hg |
| Catalyst Diameter (ring) | 4 - 8 mm |

*Actual value will depend on required removal level.

The first two of these variables are the most important since they determine the size of the reactor. Higher flow rates or removal levels require larger reactors. Pressure drop for FGT systems does not usually change for a particular process. To accommodate the higher flow rates, the reactor cross-sectional area is increased to provide a constant linear velocity. The $NO_x$ concentration will affect the $NH_3$ and dilution steam requirements, but will not affect reactor size. Both flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Since industrial boilers have fewer burners than utility boilers,[2] one burner represents a more significant fraction of the total boiler capacity. Therefore, a change in load on an industrial boiler may change these variables substantially if a burner is taken out of service.

The FGT system will have to be designed to accommodate load changes. The necessary design accomodations will have to be determined for each application after examining operating data and establishing ranges of variation. Most likely this will involve a slight overdesign of the reactor and other equipment. The process control loops used with utility boiler applications should be capable of following load changes in industrial boilers.

Space velocity is usually defined as the volume of catalyst or reactor required to treat a given flow rate of flue gas.[16] The magnitude of the space velocity is dependent entirely on the particular catalyst being considered. As can be seen in Table 2.3.2-1, the range for moving bed processes is 6000-10,000 $hr^{-1}$. These values are typically reported for 90 percent removal. For lower control levels, the value will be proportionally greater.

Almost all SCR processes require temperatures in the 350-400°C range in order to achieve good reaction rates. The temperature can vary with such things as boiler load, excess air, and ambient air temperatures. To control temperature two techniques are possible. The first involves bypassing a portion of the hot flue gas around the economizer and mixing with the economizer outlet gas so that the desired temperature is maintained.[11] The other technique uses inline heaters to obtain the desired temperature.[17]

The $NH_3:NO_x$ mole ratio is also a function of the necessary removal level and, to a lesser extent, space velocity.[18] For the three levels of removal considered in this study, 70, 80 and 90 percent, $NH_3:NO_x$ mole ratios of 0.7, 0.8, and 0.9 are required, respectively. These data are for oil-fired boilers.

The catalyst shape and size is determined by the process vendor and is simply a design decision. Ring shapes (shaped like Raschig rings) are the most resistant to particulate plugging and, for this reason, were selected for this study.

The most recently published cost estimates for SCR systems are those of the Japanese Environment Agency which were published in Dr. Jumpei Ando's most recent report on Japanese $NO_x$ control technology. Values taken from this study for two gas flow rates are shown below. The smaller gas flow rate is typical of industrial sized units while the larger flow rate is typical of utility installations.[19]

| Gas Flow Rate | Capital Cost | Operating Cost |
|---|---|---|
| 50,000 $Nm^3$/hr | $\$0.5 \times 10^6$ | $\$0.2 \times 10^6$ |
| 1,200,000 $Nm^3$/hr | $\$5.0 \times 10^6$ | $\$3.5 \times 10^6$ |

The cost for the large unit was included for comparison with other costs for large units that were reported for other process types in Section II. The reactor and catalyst type (fixed packed bed, moving bed or parallel flow) were not disclosed and, as a result, those costs are assumed to apply to all $NO_x$-only SCR systems.

Vendors of moving bed SCR systems are listed in Table 2.3.2-2 and the relative levels of application are noted. Although there are seven vendors, only four have applied their process to boilers. Of these, three have been applied to commercial scale equipment. Table 2.3.2-3 lists the moving bed systems that have been applied to oil-fired industrial boilers in Japan. Moving bed systems have not been applied to utility boilers in that country. Presently, there are no moving bed systems operating in the U.S. The Japanese installations all treat gas from residual oil-fired boilers, implying that the technology is not necessary for distillate oil-fired applications which can use fixed packed beds.

### 2.3.2.2 System Performance--

The performance of several moving bed catalysts and plants is illustrated in Figures 2.2.3-2 through 2.3.2-8. The data presented indicate that $NO_x$ control greater than 90 percent is possible through the correct selection of process design variables. Outlet $NH_3$ concentrations are also shown. These are discussed in detail in Section VI. Table 2.3.2-4 shows several operating values from a commercial installation.

TABLE 2.3.2-2. VENDORS OF SCR MOVING BED SYSTEMS FOR OIL-FIRED APPLICATIONS[21]

| Vendor | Notes |
|---|---|
| Sumitomo Chemical & Mitsubishi Heavy Industries | Tested on commercial scale equipment |
| Hitachi, Ltd. | Tested on commercial scale equipment |
| Ishikawajima-Harima Heavy Industries | Tested on pilot scale equipment |
| Kurabo | Tested on commercial scale equipment |
| Kobe Steel | Has not been applied to boilers |
| Sumitomo Heavy Industries | Has not been applied to boilers |
| Asahi Glass Company | Has not been applied to boilers |

TABLE 2.3.2-3. EXISTING FGT INSTALLATIONS OF SCR MOVING BED SYSTEMS OIL-FIRED INDUSTRIAL BOILERS[21]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3$/hr) | Completion Date |
|---|---|---|---|---|---|
| Kaizuka | Chiyoda Kenzai | Hitachi, Ltd. | Resid | 15,000 | Oct 1977 |
| Amagasaki | Nippon Oils & Fats | Hitachi, Ltd. | Resid | 20,000 | Apr 1978 |
| Sodegaura | Sumitomo Chemical | Mitsubishi H.I. | Resid | 300,000 | Sept 1976 |
| Sodegaura | Sumitomo Chemical | Sumitomo Chemical/ Mitsubishi H.I. | Resid | 300,000 | Oct 1976 |
| Hirakatu | Kurabo | Kurabo | Resid | 30,000 | Aug 1975 |

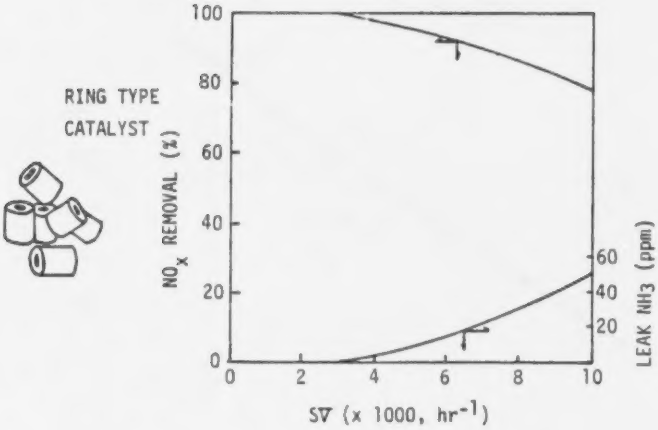


Figure 2.3.2-3. SV *vs*. $NO_x$ removal and $NH_3$ leak (ring type catalyst, 15 mm diameter, 350°C $NH_3/NO$ 1.0, inlet $NO_x$ 250 ppm).[109]

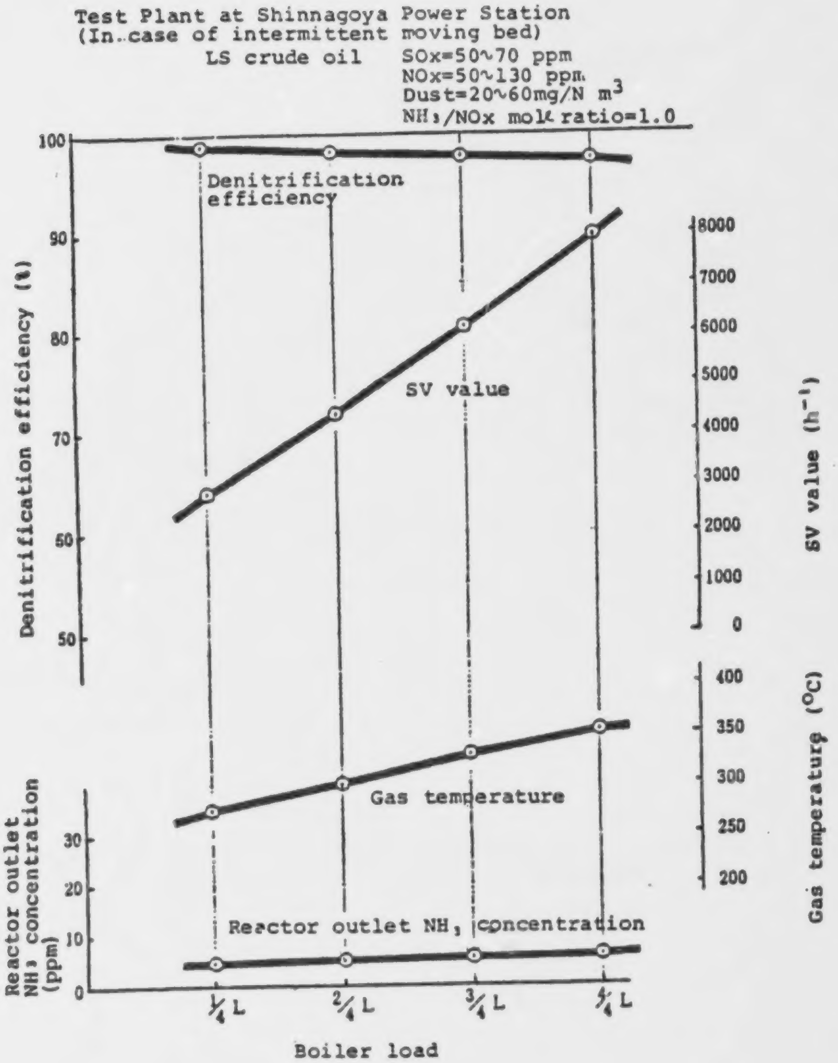


Figure 2.3.2-4. Relation between boiler load and denitrification efficiency (one example).[110]

## Intermittent Moving Bed Reactor Test Data

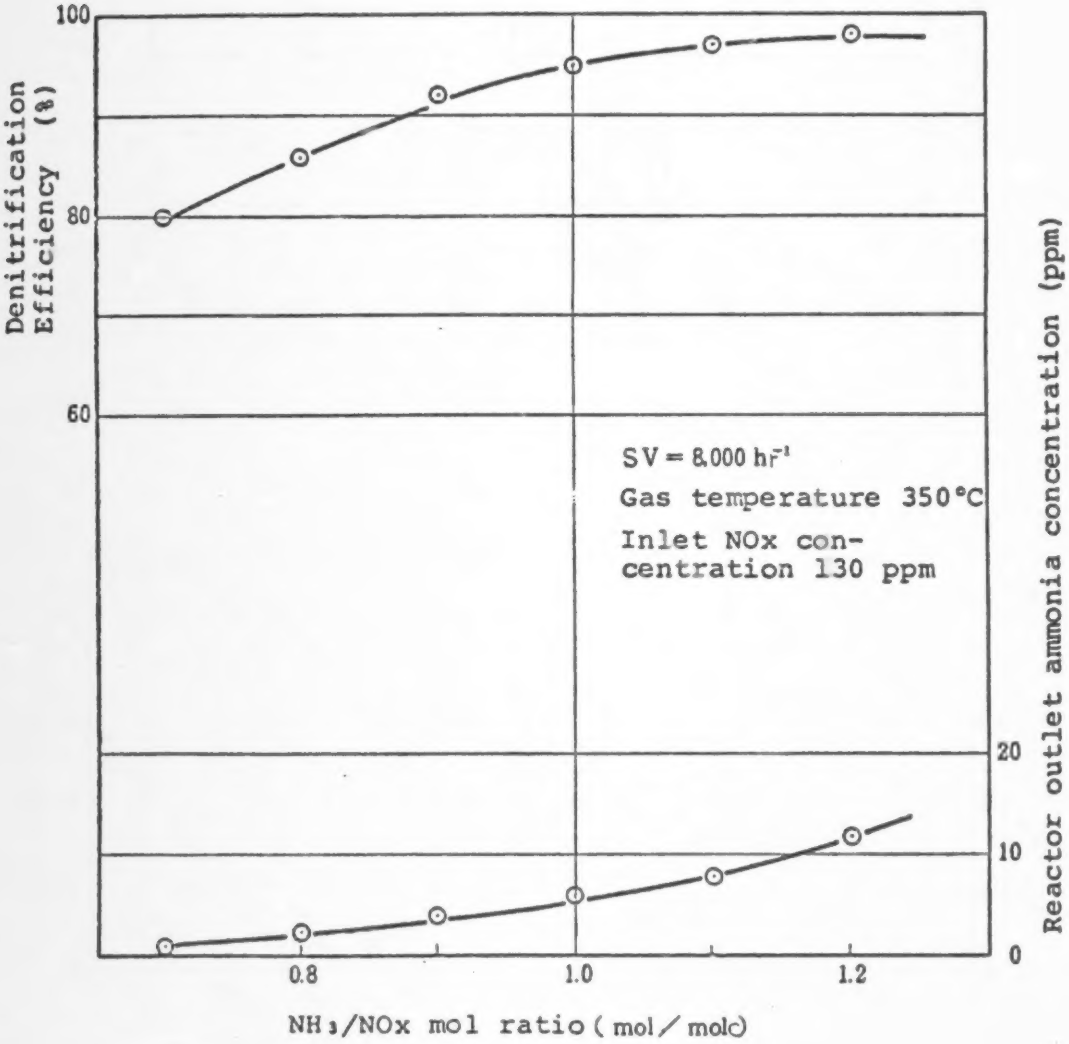


Figure 2.3.2-5. $NH_3/NO_x$ mole ratio *vs*. denitrification efficiency and reactor outlet ammonia concentration.[111]

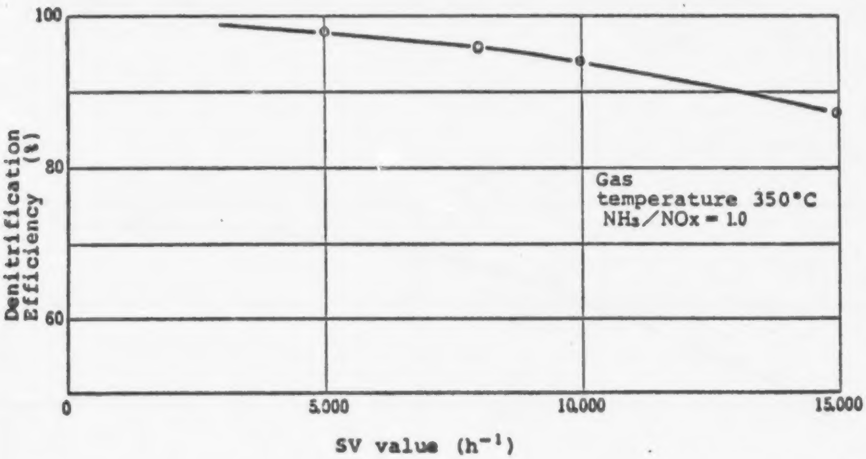
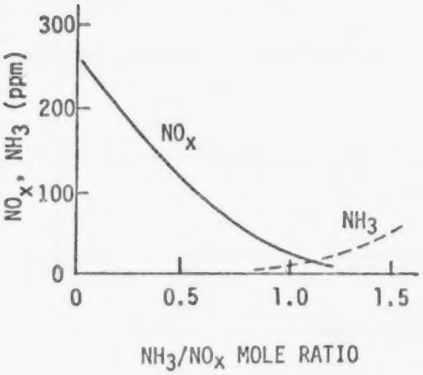


Figure 2.3.2-6. SV value and denitrification efficiency [112] (for small, <1 mm, diameter particles).



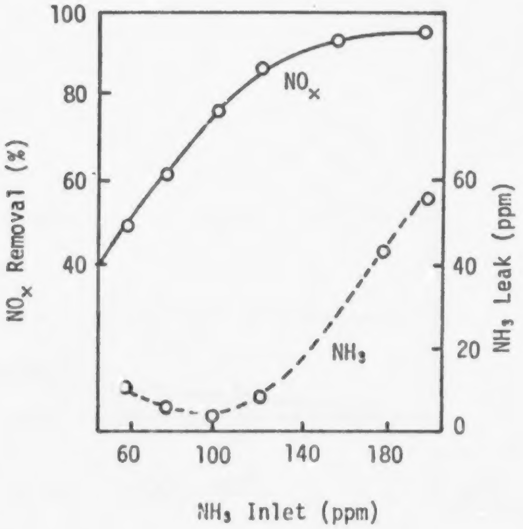


Figure 2.3.2-7. Relationship of NH$_3$/NO$_x$ ratio to outlet NO$_x$, NH$_3$ concentrations.[113]

Figure 2.3.2-8. At 300°C.[114]

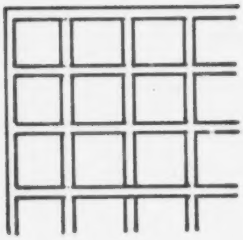
TABLE 2.3.2-4. OPERATION DATA OF A COMMERCIAL SCR PLANT FOR DIRTY GAS[108]

| | |
|---|---|
| Gas for SCR ($Nm^3$/hr) | 300,000 |
| Fuel | Oil (S=0.7%) |
| Load fluctuation | 60-100% |
| Stack height (m) | 140 |
| Inlet gas composition | |
| $O_2$ (%) | 6 |
| $SO_x$ (ppm) | 400 |
| $NO_x$ (ppm) | 200 |
| Particulates after EP (mg/$Nm^3$) | 10-20 |
| FGD unit | Scheduled |
| SV ($hr^{-1}$) | 5,000 |
| Temperature (°C) | 320 |
| $NO_x$ removal (%) | Over 90 |
| $NH_3$/NO mole ratio | 1.0 |
| Leak ammonia (ppm) | 10-20 |
| Type of reactor | Moving bed |
| Plant completed | Oct. 1976 |

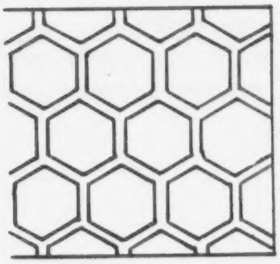
## 2.3.3 Selective Catalytic Reduction-Parallel Flow Reactor

### 2.3.3.1 System Description--

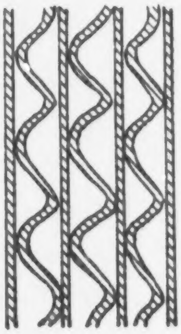
The distinguishing aspect of this process is the catalyst shape which is produced in a variety of shapes. The catalysts are produced in either a honeycomb, pipe, or plate shape. Both metal and ceramic supports are employed. Several shapes are illustrated in Figure 2.3.3-1. The catalyst shapes allow particulate laden flue gas to pass through the reactor with no inertial impaction of the particles while the $NO_x$ is transported to the catalyst surfaces by basic diffusion. The catalysts can handle all of the particulate levels emitted by the standard boilers.



Honeycomb
(Ceramic)
(Grid Type)



Honeycomb
(Ceramic)
(Hexagonal)



Honeycomb
(Metal)
(Wave Type)

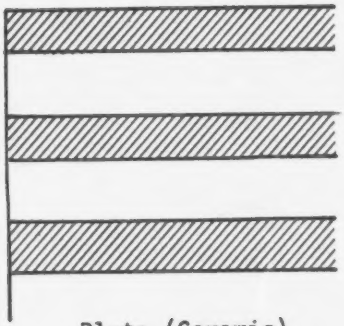


Plate (Ceramic)

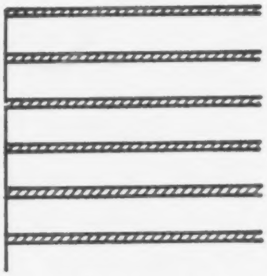
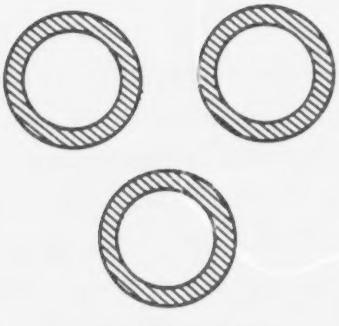
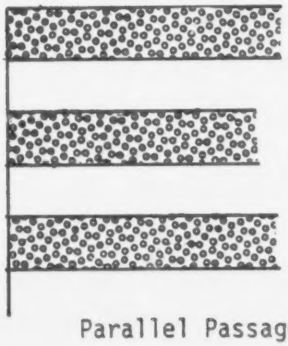


Plate (Metal)



Tube (Ceramic)



Parallel Passage

Figure 2.3.3-1. Shapes of parallel flow catalysts.[22]

The reactors used are similar to standard fixed bed units and an example is shown in Figure 2.3.3-2. The catalyst is usually prepared in small modules and manually stacked within the reactor. The specific arrangement will depend on the particular process under consideration.

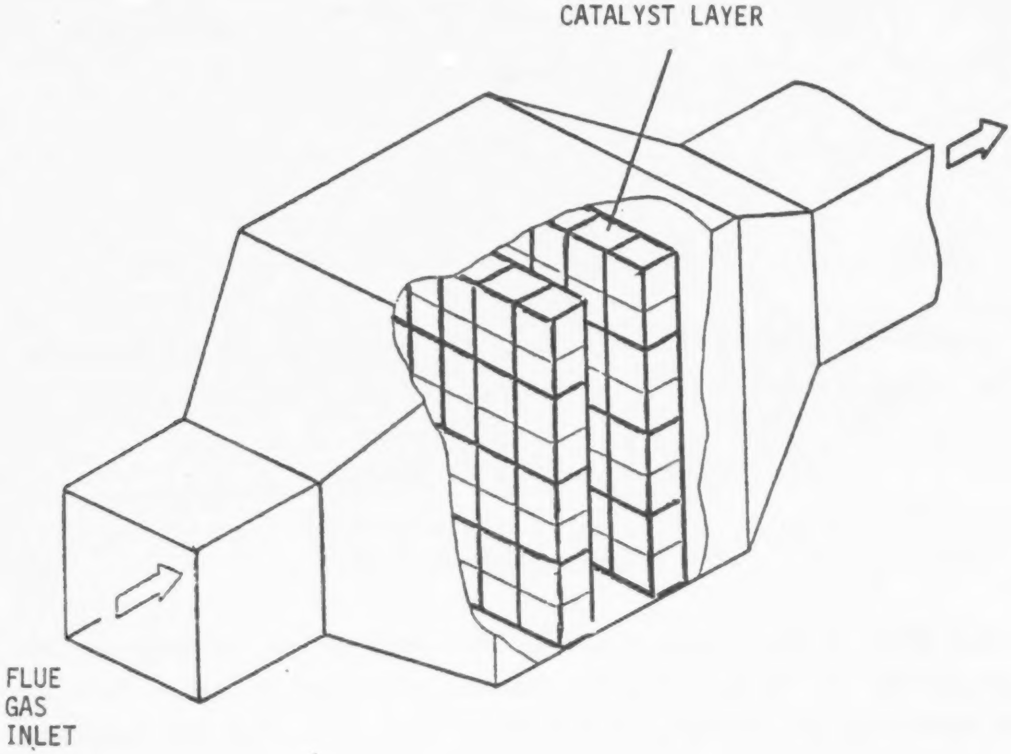


Figure 2.3.3-2. Typical reactor used with parallel flow SCR process.[23]

A typical flow diagram for a parallel flow SCR system is shown in Figure 2.3.3-3. The arrangement is similar to the other SCR processes in that hot flue gas leaving the boiler economizer is injected with $NH_3$ and passed through a catalyst bed. Temperature control is important and can be accomplished with either a fired heater or an economizer bypass. $NH_3$ can be controlled using boiler operating condition inputs to conventional control components.

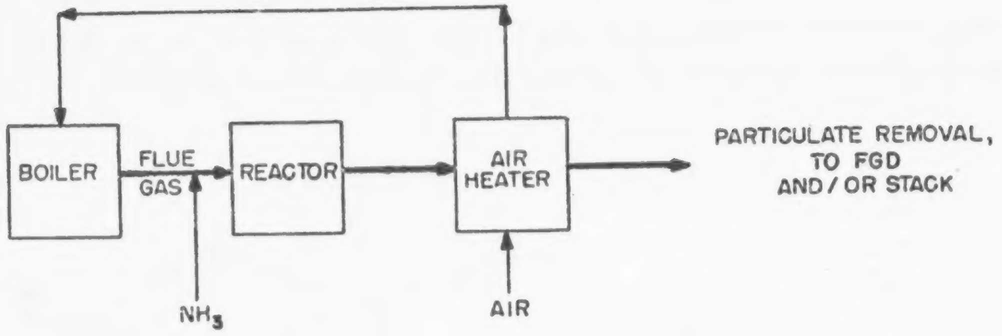

BOILER
FLUE
GAS
REACTOR
AIR
HEATER
PARTICULATE REMOVAL,
TO FGD
AND / OR STACK
NH3
AIR

Figure 2.3.3-3. Flow diagram for parallel flow SCR process.[24]

Within the reactor, $NO_x$ reacts with $NH_3$ to form $N_2$ and $H_2O$ according to the following reactions.[12]

$$4NO + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \qquad (2\text{-}1)$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \qquad (2\text{-}2)$$

Reaction (2-1) is the primary reaction since flue gas $NO_x$ is typically 90-95 percent NO. $O_2$ is necessary for both reactions and is present in sufficient quantities (>3 percent) in all of the flue gases from the standard boilers.

The catalyst volume for a desired $NO_x$ removal can be determined by the fundamental design equation for a plug flow reactor.[13]

$$\frac{V}{F} = \int_o^x \frac{dx}{r} \qquad (2\text{-}3)$$

The reaction rate, r, can be expressed as

$$r = k[NH_3]^a\ [NO]^b\ [O_2]^c \qquad (2\text{-}4)$$

The variables presented here have the same definitions as those presented in equations 2-3 and 2-4 of Section 2.2.2. The catalyst volume can also be determined if the space velocity is known for the catalyst and removal level of interest. The space velocity is defined as the gas flow rate divided by the catalyst volume.

The reaction rate is different for each catalyst formulation since different catalysts will lower the activation energy by different amounts. The activation energy affects the reaction rate constant, k, according to the Arrhenius equation.

$$k = Ae^{-\frac{E}{RT}} \qquad (2\text{-}5)$$

An important design variable with catalytic systems is the space velocity which expresses the volume of catalyst required to treat one volume per hour of flue gas. Space velocity varies with catalyst formulation, catalyst shape, and control level. Typical values of space velocity for various catalyst shapes are shown in Table 2.3.3-1. Also shown are other catalyst design variables such as catalyst dimensions, gas velocities, bed depth and pressure drop. Ranges of values are used since specific values are different for each catalyst. The values shown pertain to 90 percent $NO_x$ removal and an $NH_3/NO_x$ mole ratio of 1:1.

Both $NH_3/NO_x$ ratio and space velocity will change with removal level. The $NH_3/NO_x$ mole ratio will range from 0.7-1.0 and the space velocity will range approximately as shown in the table for control levels of 70 to 90 percent.[15]

Variables associated with the boiler can also affect the performance of these systems. These are

- flue gas flow rate
- $NO_x$ concentration
- boiler load variability

TABLE 2.3.3-1. CATALYST DESIGN VARIABLES FOR VARIOUS CATALYST SHAPES (Basis: 90% $NO_x$ removal at $NH_3/NO_x$ ratio of 1:1, 350-400°C)

|  | Honeycomb (metallic) | Honeycomb, tube (ceramic) | Parallel Plate (Ceramic) | Parallel Plate (Metallic) |
|---|---|---|---|---|
| Catalyst size (mm) |  |  |  |  |
| Thickness | 0.5-1 | 1.5-3 | 8-10 | 1 |
| Opening | 4-8 | 6-20 | 8-14 | 5-10 |
| Gas velocity (m/sec)[a] | 2-6 | 5-10 | 5-10 | 4-8 |
| Bed depth (m) | 1-2 | 1.5-5 | 4-6 | 2-5 |
| SV (1,000 $hr^{-1}$)[b] | 5-8 | 4-8 | 1.5-3 | 2-5 |
| Pressure drop ($mmH_2O$) | 30-80 | 40-160 | 80-160 | 60-120 |

[a] Velocity at 350-400°C in open column (superficial velocity).

[b] Gas volume ($Nm^3$/hr)/catalyst bed volume ($m^3$).

The flue gas flow rate and control level determine the catalyst volume (hence reactor size) necessary. Increases in either also increase the reactor size. The $NO_x$ concentration is a function only of fuel type used in the standard boilers. Higher concentrations require larger $NH_3$ storage and vaporization equipment; reactor size is not affected. Boiler load can affect several things including flue gas temperature, flow rate and $NO_x$ concentration. It is necessary to maintain reactions temperatures of 350 to 400°C and temperature control equipment may be necessary if the boiler experiences large load variations. Where these variations are present, some equipment overdesign may be warranted to insure a constant control level. These variables are discussed in more detail in the section on moving bed SCR systems for oil-fired boilers, Section 2.3.2.

Parallel flow SCR processes have been applied in Japan to several residual oil-fired industrial boilers. Oil-fired utility boilers and other sources with high particulate concentrations are also being treated. A list of vendors of parallel flow SCR systems is presented in Table 2.3.3-2. Notes on the relative level of application are also shown. Four of the eight vendors have applied their systems to oil-fired boilers indicating that application of this technology to industrial boilers is technically feasible. Parallel flow SCR systems have been applied to both industrial and utility boilers. Specific applications are listed in Tables 2.3.3-3 and 2.3.3-4. There have been no applications in the U.S. The tables indicate that the parallel flow technology is designed primarily for residual oils and not distillate oils.

TABLE 2.3.3-2. VENDORS OF SCR PARALLEL FLOW SYSTEMS FOR OIL-FIRED APPLICATIONS[21]

| Vendor | Notes |
|---|---|
| Hitachi Zosen | Tested on pilot scale equipment |
| Hitachi, Ltd. | Tested on commercial scale equipment |
| JGC | Has not been tested on boilers |
| Mitsui Engineering & Shipbuilding | Tested on commercial scale equipment |
| Mitsubishi Heavy Industries | Tested on commercial scale equipment |
| Ishikawajima-Harima Heavy Industries | Tested on commercial scale equipment |
| Kobe Steel | Has not been tested on boilers |
| Kawasaki Heavy Industries | Tested on pilot scale equipment |

TABLE 2.3.3-3. EXISTING FGT INSTALLATIONS OF SCR PARALLEL FLOW SYSTEMS OIL-FIRED INDUSTRIAL BOILERS[21]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Sodegaura | Fuji Oil | Mitsubishi H.I. | Resid | 200,000 | January 1978 |
| Kawasaki | Ajinomoto | Ishikawajima H.I. | Resid | 180,000 | January 1978 |
| Chiba | Ukishima Pet. Chem. | Mitsui Engineering | Resid | 220,000 | April 1978 |

TABLE 2.3.3-4. EXISTING FGT INSTALLATIONS OF SCR PARALLEL FLOW SYSTEMS OIL-FIRED UTILITY BOILERS[21]

| Location (Japan) | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Yokosuka | Tokyo Electric | Mitsubishi H.I. | Resid | 40,000 | March 1977 |
| Chita | Chubu Electric | Mitsubishi H.I. | Resid | 1,920,000 | February 1980 |
| Kudamatsu | Chugoku Electric | Ishikawajima H.I. | Resid | 1,900,000 | July 1979 |
| Niigata | Tohoku Electric | Ishikawajima H.I. | Resid | 1,660,000 | August 1981 |

### 2.3.3.2 System Performance--

The performance of several parallel flow catalysts is illustrated in Figures 2.3.3-4 through 2.3.3-9. Table 2.3.3-5 shows several operating data for a single parallel flow SCR installation. The data presented indicated that $NO_x$ control levels of greater than 90 percent are obtainable through selection of the appropriate process design variables. Other data are also presented and these are discussed in subsequent sections.

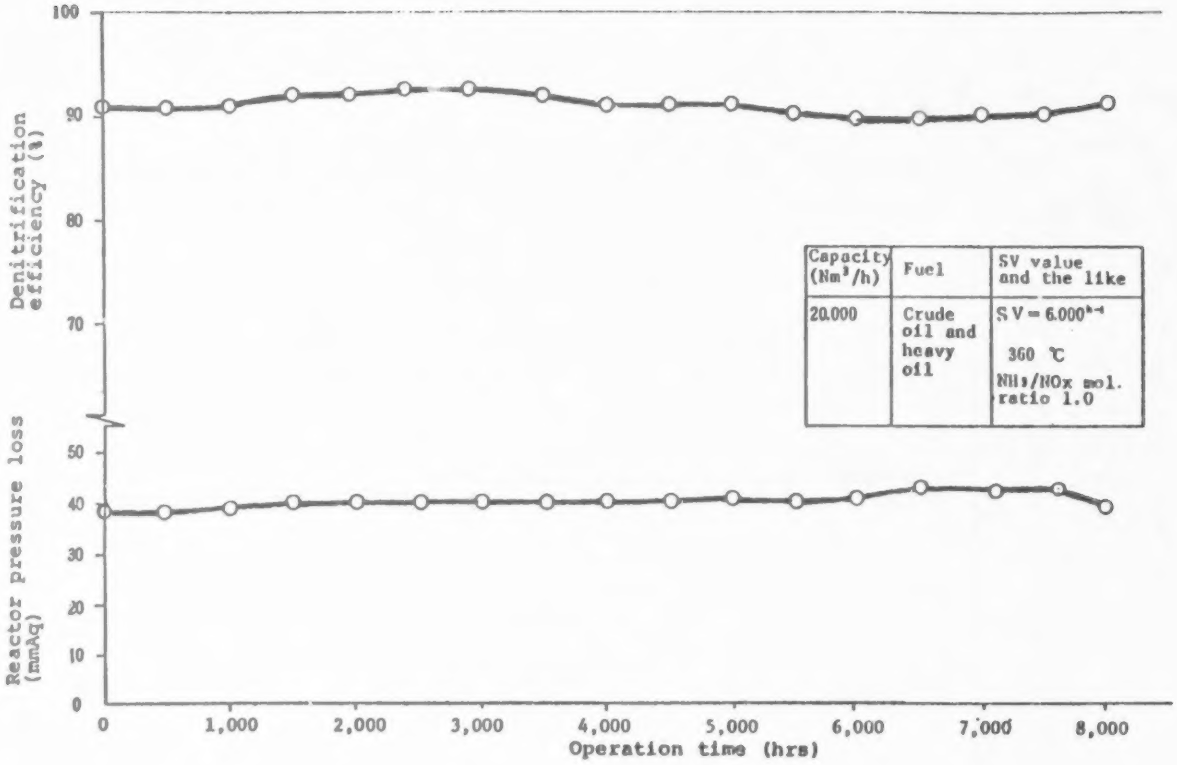


Figure 2.3.3-4. Catalyst life test results.[115]

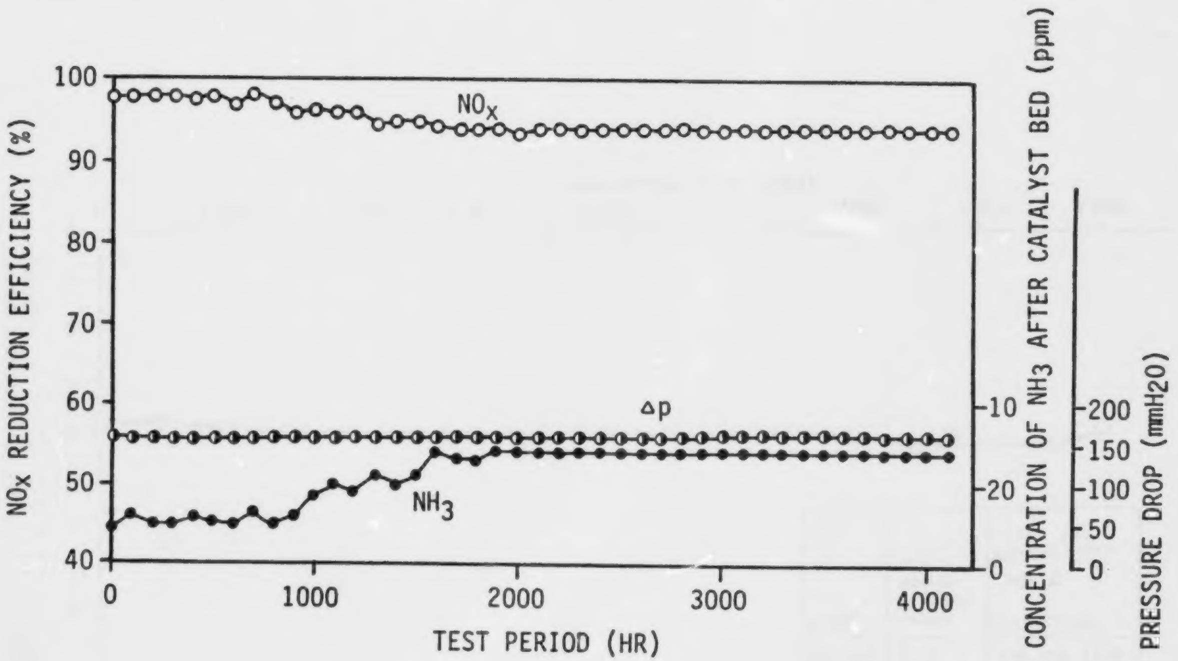


Figure 2.3.3-5. Durability of $NO_x$ removal catalyst for exhaust gas of high sulfur oil burning boiler.[116]

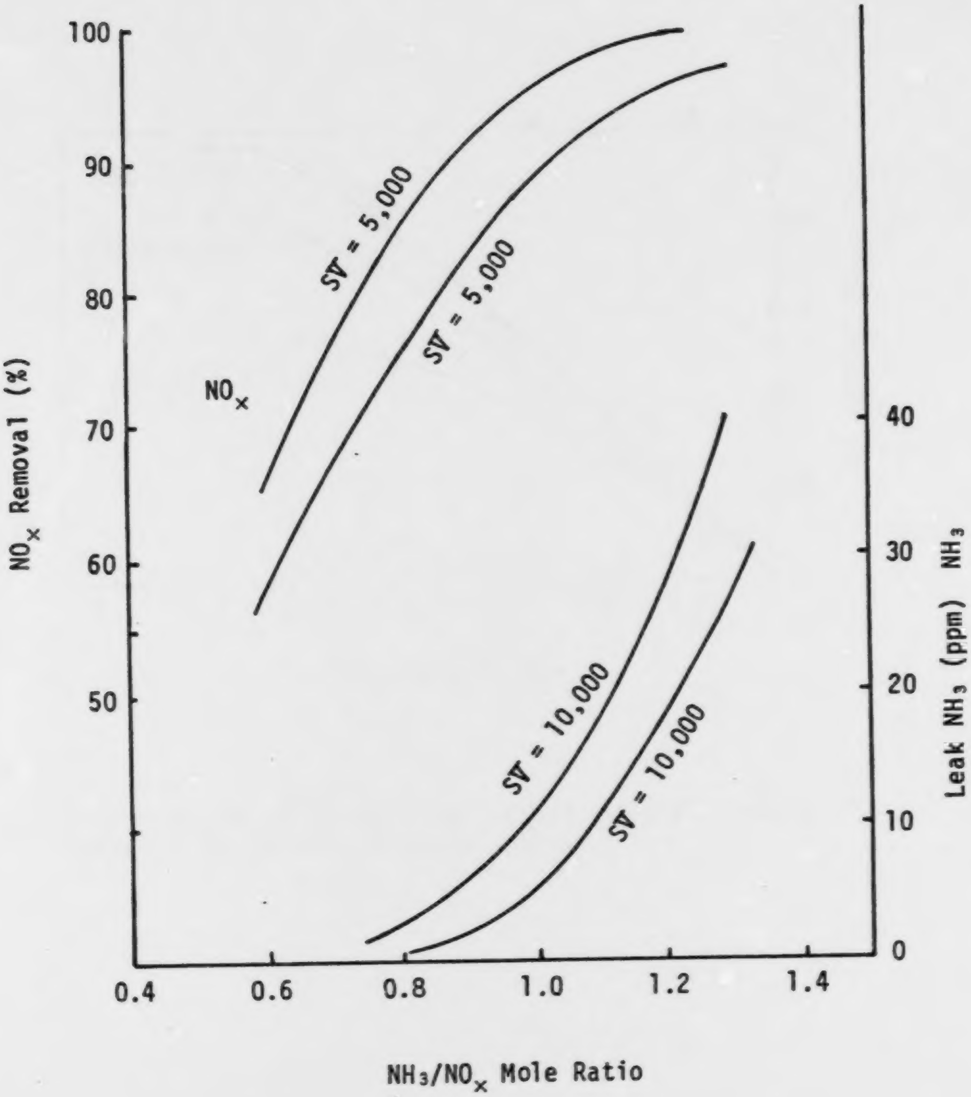


Figure 2.3.3-6. Typical example of operation data (oil-fired boiler, 350-400°C, granular or honeycomb catalyst).[105]

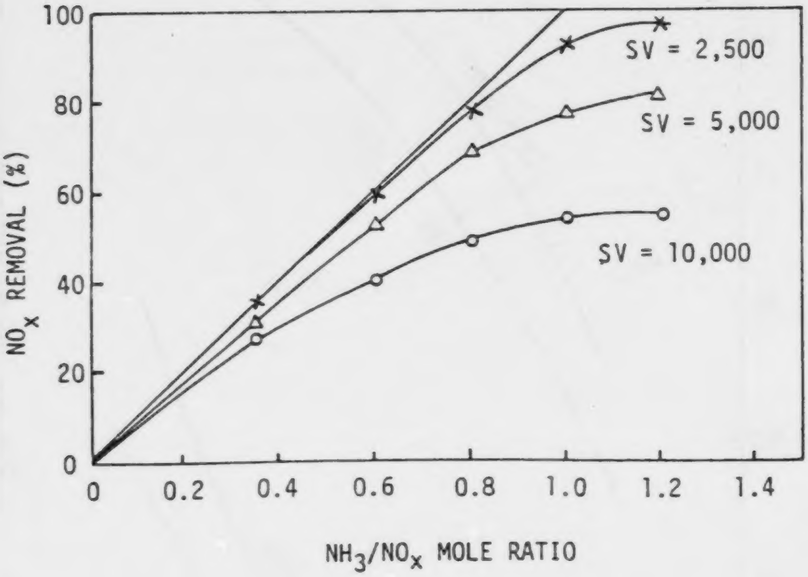


Figure 2.3.3-7. $NH_3/NO_x$ mole ratio *vs.* $NO_x$ removal (plate catalyst; 350°C, LV 5.9 m/sec).[117]

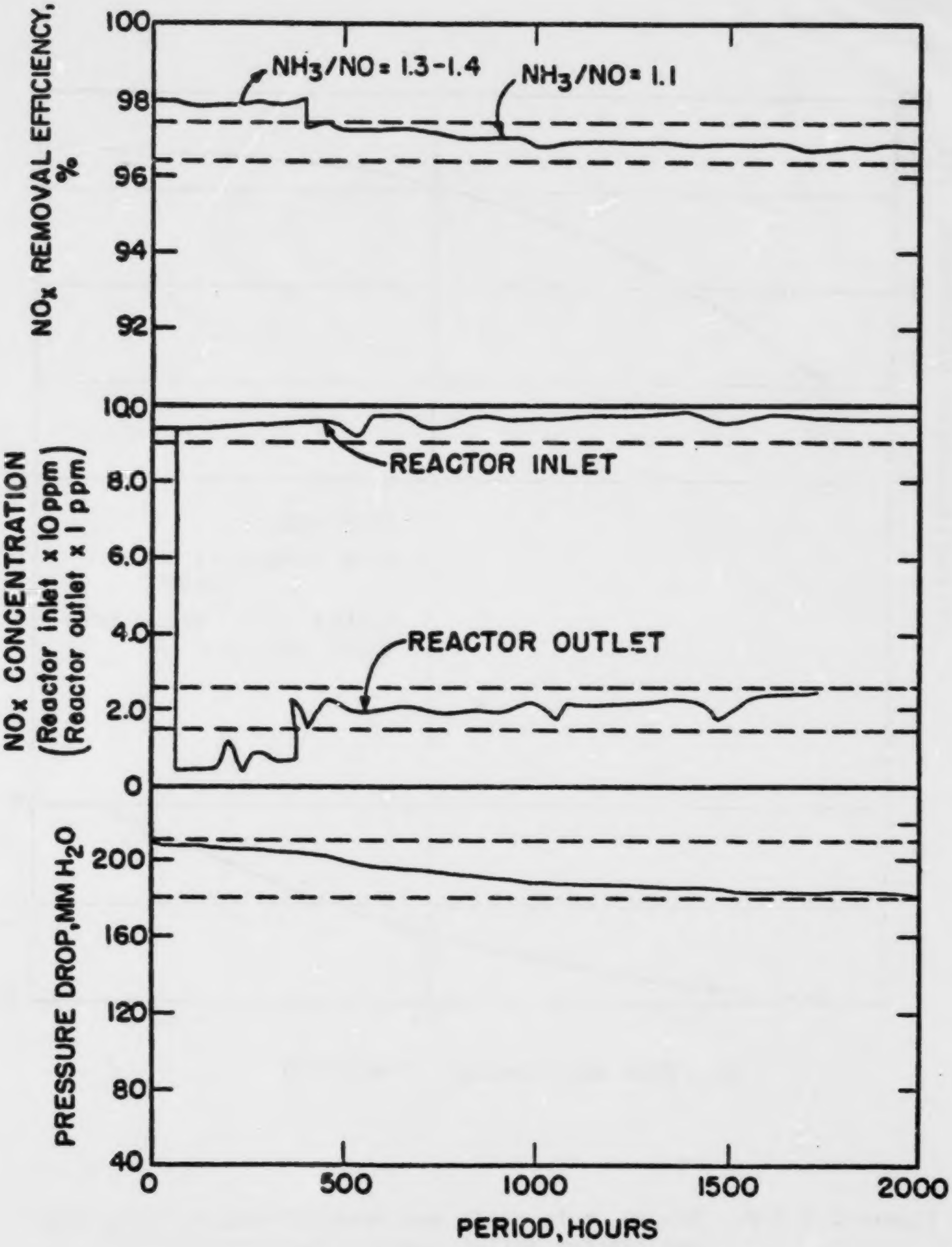


Figure 2.3.3-8. $NO_x$ removal efficiency, $NO_x$ concentration, and pressure loss over 2,000 hr test period for JGC Paranox Process.[118]

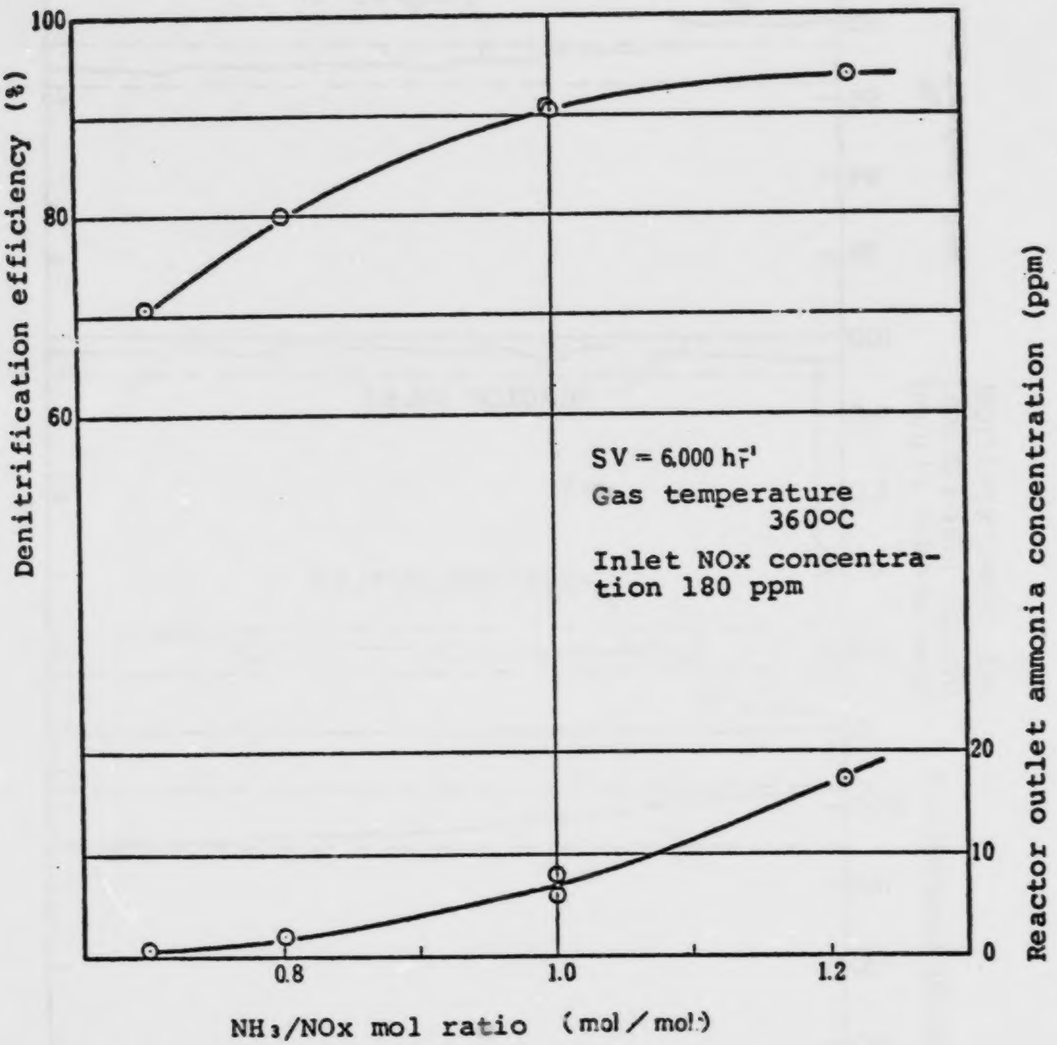


Figure 2.3.3-9. $NH_3/NO_x$ mole ratio and denitrification efficiency and reactor outlet ammonia concentration.[119]

TABLE 2.3.3-5. SCR PLANT BY MITSUI ENGINEERING AND SHIPBUILDING CO.[20]

| | |
|---|---|
| Capacity ($Nm^3/hr$) | 220,000 |
| Gas composition | |
| $NO_x$ (ppm) | 150 |
| $SO_x$ (ppm) | 300 |
| Dust ($mg/Nm^3$) | 100-150 |
| Catalyst and reactor | |
| Catalyst carrier | TiO |
| Catalyst shape | PP |
| SV ($hr^{-1}$) | 4,000 |
| Temperature (°C) | 350-400 |
| $NH_3/NO_x$ mole ratio | 1.0 |
| $NO_x$ removal (%) | Above 90 |
| Total pressure drop ($mmH_2O$) | 180 |
| Leak $NH_3$ (ppm) | Below 10 |
| Operation start | July 1977 |
| Plant cost ($10^6$ yen) | 260 |

### 2.3.4 Absorption-Oxidation

#### 2.3.4.1 System Description--

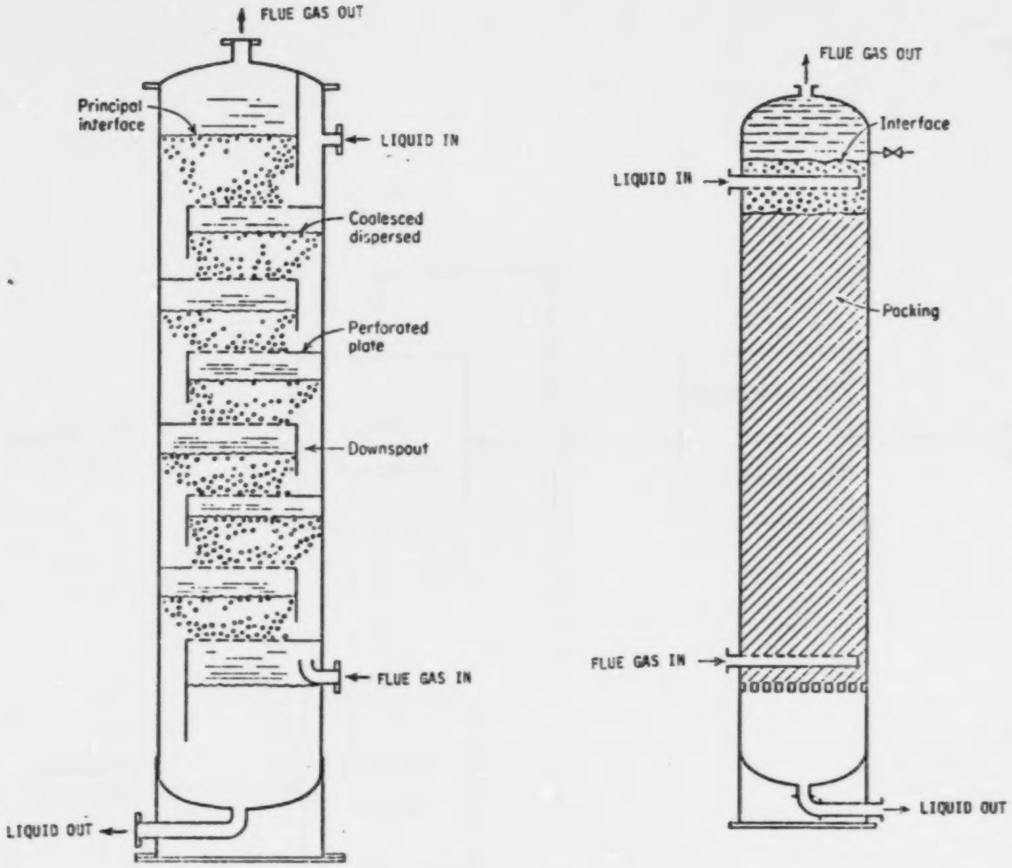
**Absorption-oxidation processes remove $NO_x$ from flue gas by absorbing the NO or $NO_2$ into a solution containing an oxidant which converts the $NO_x$ to a nitrate salt. Two types of gas/liquid contactors can be used and examples of each type are shown in Figure 2.3.4-1. Both perforated plate and packed towers accomplish $NO_x$ absorption by generating high gas/liquid interfacial areas. The choice of one type of contactor is a design decision made to achieve a given removal for the least cost.**

A generalized process flow diagram is shown in Figure 2.3.4-2. Flue gas is taken from the boiler after the air preheater. Before the gas can be sent to the $NO_x$ absorber, it must be $SO_2$-free since $SO_2$ consumes prohibitive amounts of the costly liquid-phase oxidant. In most cases, the oxidant is permanganate ($MnO_4^-$); however, $Ca(ClO)_2$ can also be used. Therefore, a conventional FGD unit is required ahead of the $NO_x$ absorber. A prescrubber to cool the gas and remove both particulates and $Cl^-$ prior to FGD is also necessary. After having passed through these two scrubbing sections, the flue gas enters the distributing space at the bottom of the $NO_x$ absorber, below the packing or plates. The gas passes upward through the column, countercurrent to the flow of the liquid absorbent/oxidant (usually a KOH solution containing $KMnO_4$). $NO_x$ is absorbed and then oxidized over the length of the column according to the following reactions.[31]

$$NO(g) \rightarrow NO(aq) \qquad (2\text{-}6)$$

$$NO(aq) + KMnO_4(aq) \rightarrow KNO_3(aq) + MnO_2(s) \qquad (2\text{-}7)$$

$$2NO_2(g) \rightarrow N_2O_4(g) \qquad (2\text{-}8)$$


FLUE GAS OUT
Principal interface
LIQUID IN
Coalesced dispersed
Perforated plate
Downspout
FLUE GAS IN
LIQUID OUT
FLUE GAS OUT
Interface
LIQUID IN
Packing
FLUE GAS IN
LIQUID OUT

Perforated Plate Absorber

Packed Absorber

Figure 2.3.4-1. Gas/liquid contactor options for Absorption-Oxidation Processes.[29]

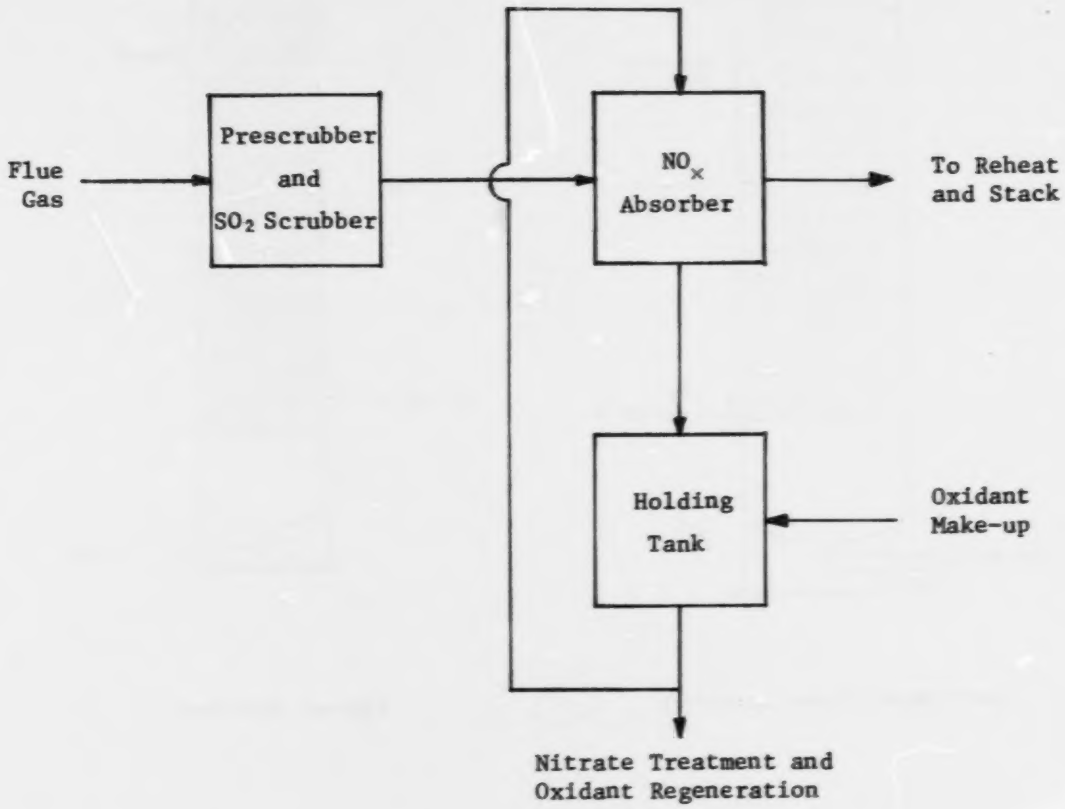

Flue
Gas
Prescrubber
and
$SO_2$ Scrubber
$NO_x$
Absorber
To Reheat
and Stack
Holding
Tank
Oxidant
Make-up
Nitrate Treatment and
Oxidant Regeneration

Figure 2.3.4-2. Process flow diagram for absorption-oxidation process.[30]

$$N_2O_4(g) \rightarrow N_2O_4(aq) \qquad (2\text{-}9)$$

$$N_2O_4(aq) + 2K_2MnO_4(aq) \rightarrow 2KMnO_4(aq) + 2KNO_2(aq) \qquad (2\text{-}10)$$

Since most of the $NO_x$ from combustion processes occurs as NO,[32] reactions 2-6 and 2-7 predominate. The clean gas passes out of the top of the absorber to a heater for plume buoyancy and is sent to the stack. The absorbing solution drops to a holding tank where makeup KOH and/or $KMnO_4$ are added. This solution flows to a centrifuge to separate the solid $MnO_2$ which is then electrolytically oxidized to $MnO_4$. The remaining solution is either concentrated in an evaporator to form a weak $KNO_3$ solution or is electrochemically treated to produce a weak $HNO_3$ solution and a mixed stream of KOH and $KNO_3$.

The fundamental design equation used for gas absorption column design is

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \qquad (2\text{-}11)$$

where y = bulk $NO_x$ concentration (mole fraction) of gas phase at any given point in column

$y-y^*$ = overall driving force for absorption ($y^*$ being the $NO_x$ concentration of a gas in equilibrium with a given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/ $(ft^2)$(hr)(mole fraction)

$a$ = area of gas-liquid interface per unit packed volume, $ft^2/ft^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/$(ft^2)(hr)$

$Z$ = length of packed section of column, ft

In a column containing a given packing or plate configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity, obviously, must be lower than the flooding velocity. How much lower is a design decision. Often, it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[34] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter is a major problem for these systems trying to achieve large $NO_x$ reductions since NO is relatively insoluble in water. This can be seen in Table 2.3.4-1.

TABLE 2.3.4-1. NITROGEN OXIDES CHARACTERISTICS[35]

| | Boiling Point, °C | Solubility in Cold Water (0°C), $cm^3$ | Solubility in Hot Water (60°C), $cm^3$ |
|---|---|---|---|
| NO | -151.8 | 7.34/100 cc $H_2O$ | 2.37/100 cc $H_2O$ |
| $NO_2$ | 21.2 | soluble, decomposes | -- |

One can see that NO has a very limited solubility in water and, since most $NO_x$ is present as NO, the rate of mass transfer (absorption) is going to be relatively slow. This means that the absorber must be tall with a high liquid flow rate. Table 2.3.4-2 presents the effects of boiler/flue gas variables on the design of absorption-oxidation systems.

TABLE 2.3.4-2. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Requires FGD pretreatment |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased oxidant concentration |

Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load changes. The column size and the liquid and oxidant flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

None of the sources consulted for this study could supply typical ranges for operating variables such as liquid/gas ratio, reagent concentrations or pressure drops and, as a result, none are presented here. Economic data were not presented either. One source did estimate the removal for absorption-oxidation processes to be 85 percent.[36]

Presently, absorption-oxidation processes are still in the pilot unit stage of development. Table 2.3.4-3 presents a list of absorption-oxidation process vendors and the status of development of their projects.

TABLE 2.3.4-3. PROCESS VENDORS OF ABSORPTION-OXIDATION PROCESSES[37,38]

| Vendor | Status of Development |
|---|---|
| Hodogaya | No information available; stopped development on process |
| Kobe Steel | 1974: 1000 $Nm^3$/hr gas from iron-ore sintering furnace; stopped development on process |
| MON (Mitsubishi Metal, MKK, Nikon Chemical) | 1974: 4000 $Nm^3$/hr flue gas from oil-fired boiler |
| Nissan Engineering | 1972: 4 pilot plants, 100-2000 $Nm^3$/hr tail gas from $HNO_3$ plant |

Only one of the process vendors has piloted this process on flue gas from an oil-fired boiler and there have been no commercial applications either in Japan or the U.S.

2.3.4.2 System Performance--

No information has been published on tests conducted with flue gas from oil-fired boilers. The relative insolubility of NO in water may present a major obstacle to achieving the stringent level of control (90 percent $NO_x$ reduction) by absorption-oxidation processes. Another primary drawback of absorption-oxidation systems is the production of nitrate salts (see Equation 2-7), a secondary pollutant. These processes probably could not be applied on a large scale as wastewater treatment systems (chemical or biological) do not remove nitrogen compounds from the wastewater.[39] Trying to recover the nitrates as nitric acid for industrial use or potassium nitrate for fertilizer does not seem promising as the by-products are of low quality. Also, the use of an expensive, liquid-phase oxidant requires stainless steel and other corrosion resistant materials of construction. High sulfur coals require an FGD system prior to the $NO_x$ absorber to prevent excessive oxidant consumption by $SO_2$. The process steps of several absorber columns in series (large fan requirements), oxidant regeneration (electrolysis), and flue gas reheat (inline heater) are all energy intensive and present technical and economic disadvantages.

### 2.3.5 Selective Catalytic Reduction-$NO_x$/$SO_2$ Removal

#### 2.3.5.1 System Description--

From a $NO_x$ removal standpoint, this process is very similar to those discussed in Sections 2.3.1 through 2.3.3. The primary difference is the additional equipment necessary to collect and process the $SO_2$. The main feature of the process is the reactor and catalyst which remove both NO and $SO_2$. This process was developed by Shell although the U.S. licensor, UOP, is currently marketing and developing the process. The $NO_x$/$SO_2$ version of the process is commonly called the SFGT process which stands for the Shell Flue Gas Treatment Process.

A uniquely designed parallel flow type of reactor is used to avoid problems with particulates. This design is necessary only with flue gas from residual oil-fired boilers. The reactor consists of a series of packages containing catalyst material, arranged in a parallel configuration which allows flue gas flow between the packages. Each package consists of catalyst material placed between two layers of wire gauze. Figure 2.3.5-1 illustrates the internals of the parallel passage reactor. The flue gas flows between the catalyst packages and not directly through the catalyst material. This prevents plugging of the catalyst with particulate matter in the flue gas.

For convenient fabrication and handling, catalyst packages of a standard size are appropriately spaced and placed in a container to form a unit cell or module. $SO_2$ removal efficiency and capacity are determined by the number of unit cells placed in series in a cell stack. For a given level of $SO_2$ removal, a greater number of cells in the stack increases the capacity and reduces the frequency of regeneration. The number of stacks is determined largely by the flue gas rate and the flue gas velocity through a single stack is generally not a design variable. For most design situations, 4 to 5 unit cells in a stack are adequate to achieve high $SO_2$ removal, however, if a high level of denitrification is required, more unit cells per stack may be necessary.

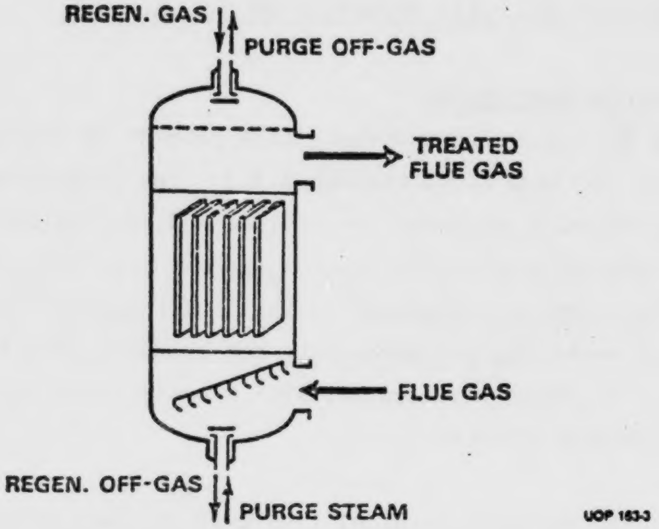


Figure 2.3.5-1. The SFGT parallel flow reactor.[40]

The SFGT process is a dry process with two or more reactors operating in a cyclic manner. The desulfurization aspect of the process is regenerable, while $NO_x$ removal is accomplished by catalytic reduction with ammonia. The catalyst material is commonly called an acceptor since $SO_2$ removal involves adsorption or "acceptance" of $SO_2$. The desulfurization cycle consists of the following steps:

1) oxidation of acceptor bed/acceptance of $SO_2$,
2) purge reactor,
3) regeneration with reducing gas, and
4) purge reactor.

The products of the oxidation and acceptance reactions in step 1 above catalyze the reaction of $NO_x$ with ammonia to form nitrogen and water. $NO_x$ removal is accomplished by metering ammonia into the untreated flue gas upstream of the reactors. The catalytic reaction takes place across the partially spent acceptor beds.

Also associated with the SFGT process are facilities for generating reducing gas and for the processing of $SO_2$ in regeneration off gases into sulfur by-products. Figure 2.3.5-2 illustrates the process flow for a typical SFGT system.

Boiler flue gas is withdrawn upstream of the air preheater and particulate removal device by the SFGT system fan and discharged to the reactor inlets. The flue gas then flows through fixed bed reactors in open channels alongside and in contact with the acceptor material. Ammonia is added to the flue gas upstream of the SFGT system fan to insure complete mixing before the flue gas enters the reactor.

Fresh acceptor material is elemental copper on an alumina support. This is converted to the oxide form by flue gas oxygen shortly after initiation of the acceptance cycle. $SO_2$ is removed by reaction with the copper oxide and oxygen as the flue gas flows through the channels, converting the acceptor material to copper sulfate. Simultaneous with the desulfurization process, the reduction of flue gas $NO_x$ by ammonia is selectively catalyzed by copper oxide and copper sulfate in the acceptor bed. As the flue gas leaves the SFGT system reactors it is returned to the boiler flue gas duct downstream of SFGT fan suction.

Flue gas is fed to a reactor until an unacceptable amount of $SO_2$ begins to pass through the reactor. This occurs when a large fraction of the acceptor has been converted to the sulfate form. Flue gas flow is then diverted to another reactor and the spent reactor is isolated. Any flue gas remaining in the spent reactor is purged with an inert gas such as steam, and the regeneration cycle is initiated.

Regeneration is accomplished by passing a reducing gas through the bed countercurrent to the direction of the flue gas flow. The reducing gas, which is primarily hydrogen, reacts with the copper sulfate in the spent reactor to convert it to elemental copper. An off gas of $SO_2$ and water is

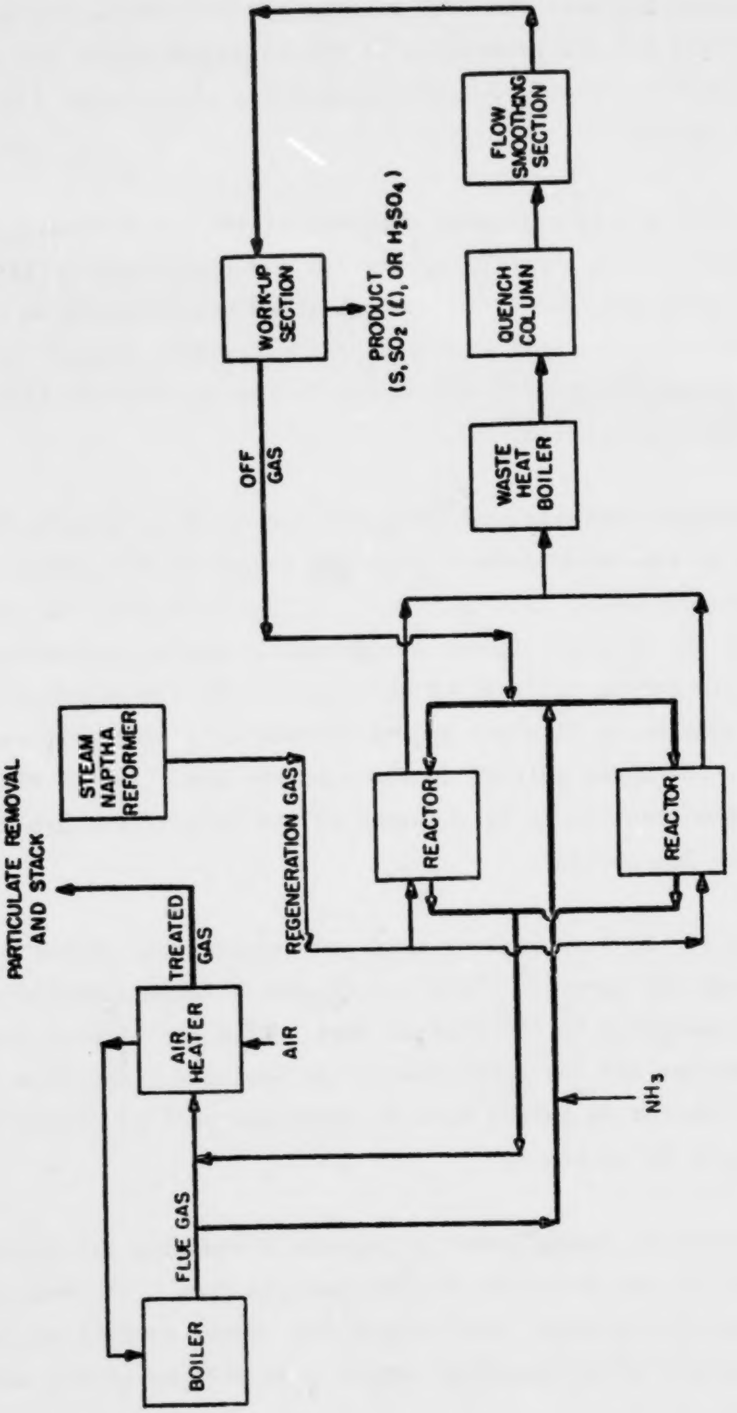

PARTICULATE REMOVAL AND STACK
STEAM NAPTHA REFORMER
TREATED GAS
AIR HEATER
AIR
FLUE GAS
BOILER
REGENERATION GAS
REACTOR
REACTOR
NH$_3$
OFF GAS
WORK-UP SECTION
PRODUCT (S, SO$_2$ (ℓ), OR H$_2$SO$_4$)
WASTE HEAT BOILER
QUENCH COLUMN
FLOW SMOOTHING SECTION

Figure 2.3.5-2. Flow diagram of the SFGT process.[41]

produced by the reaction. After regeneration is complete, the reactor is again purged with steam and is ready for another acceptance cycle. Regeneration gas can be produced from a number of sources, but steam-naphtha reforming is proposed by UOP as being the most economical.[42]

The regeneration off-gas treatment section consists of flow smoothing equipment and $SO_2$ workup equipment. Typically, the regeneration off-gas is cooled and most of the steam condensed, raising the $SO_2$ concentration from 10 percent to 80 percent by volume. The concentrated $SO_2$ is then compressed into an intermediate holding vessel to provide a smooth flow rate to the workup section. The workup section may be a modified Claus unit which produces an elemental sulfur by-product, a fractionation unit which produces liquid $SO_2$, or a sulfuric acid plant.

Each process step consists of different chemical reactions. The acceptor material is converted to the oxide form by the following reaction:

$$Cu + \tfrac{1}{2}O_2 \rightarrow CuO \qquad (2\text{-}12)$$

This oxide readily reacts with flue gas $SO_2$ and oxygen, as described by:

$$CuO + \tfrac{1}{2}O_2 + SO_2 \rightarrow CuSO_4 \qquad (2\text{-}13)$$

$SO_3$ in the flue gas is also removed by the following reaction:

$$CuO + SO_3 \rightarrow CuSO_4 \qquad (2\text{-}14)$$

The reaction scheme for reduction of $NO_x$ is described by the following:[12]

$$4NO + 4NH_3 + O_2 \rightleftarrows 4N_2 + 6H_2O \qquad (2\text{-}1)$$

$$2NO_2 + 4NH_3 + O_2 \rightleftarrows 3N_2 + 6H_2O \qquad (2\text{-}2)$$

Excess ammonia which is not consumed in reactions 2-1 and 2-2 may be catalytically oxidized to nitrogen and water by reaction with flue gas oxygen, as described by:

$$4NH_3 + 3O_2 \rightarrow 2N_2 + 6H_2O \qquad (2\text{-}15)$$

Maximum $NO_x$ removal efficiency is achieved at the point of $SO_2$ breakthrough, where conversion of the acceptor material from the oxide to the sulfate form is essentially complete. Figure 2.3.5-3 illustrates reactor outlet $SO_2$ and NO concentrations during a typical SFGT acceptance cycle.

A different set of reactions is involved during regeneration of the catalyst bed.

Copper sulfate is reduced to the elemental copper form by reducing gas hydrogen as described by the following reaction:

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O \qquad (2\text{-}16)$$

Any acceptor material present in the reactor as the oxide will also be reduced, according to the following reaction:

$$CuO + H_2 \rightarrow Cu + H_2O \qquad (2\text{-}17)$$

The regeneration step occurs at the same temperature as the acceptance step, 400°C (750°F).

The general reactor design equation is the same as that described in earlier sections for SCR processes. The primary variables are the gas rate, reaction rate, and control level. Reaction rate data have not been released for this process except that the $NO_x$ reduction is first order.

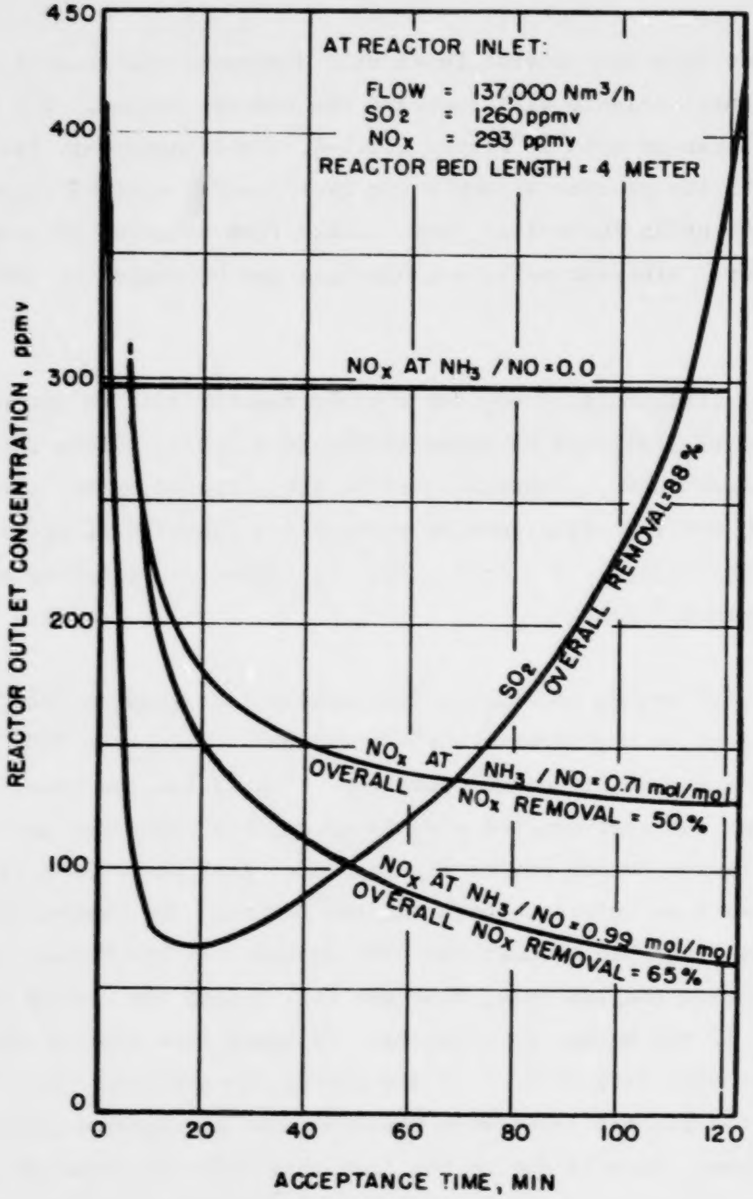


Figure 2.3.5-3. SFGT reactor performance *versus* acceptance time.[44]

The gas flow rate and control level will determine the reactor size. Increases in either variable will increase the reactor volume. The effect of control level can be seen in Figure 2.3.5-4. It is necessary for the flue gas to enter the reactor at 400°C and therefore it must be taken from an appropriate point in the boiler, most likely from between the economizer and air preheater. Alternatively, a cooler gas can be heated to 400°C by an inline heater.

The removal efficiency of $NO_x$ for a given reactor size is determined by the amount of $NH_3$ injected as shown in Figure 2.3.5-5. Since the reaction is first order in $NO_x$, control level is not affected by $NO_x$ concentration.[47] The $SO_2$ control efficiency is primarily a function of the acceptance time of the reactor (Figure 2.3.5-3). Typical ranges of operating variables are shown in Table 2.3.5-1.

Since the SFGT system can handle full particulate loading ($\geq 10$ gr/sft$^3$) it is not dependent on any pretreatment facilities. Also, the SFGT system operation is independent of boiler operation. The system fan takes suction from the flue gas duct between the economizer and air preheater and the reactor discharge returns to the boiler flue gas duct just downstream of the suction point, with no valves between the two points. The system fan provides a constant flow rate through the SFGT system. If the boiler flue gas rate is greater than the fan rate, flue gas will bypass the system through the open duct. If the boiler flue gas rate is lower than that of the system fan, treated gas will recycle back to the system fan suction. Recycle of treated gas to the reactor inlet with "open bypass" arrangement presents no operating problems. This is due to the fact that both the level of desulfurization and denitrification are independent of inlet concentrations, and the system does not humidify the flue gas.

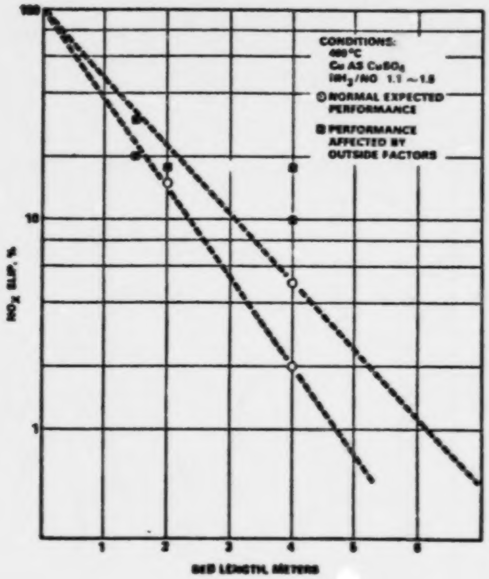


Figure 2.3.5-4. Unconverted $NO_x$ as a function of catalyst bed length.[45]

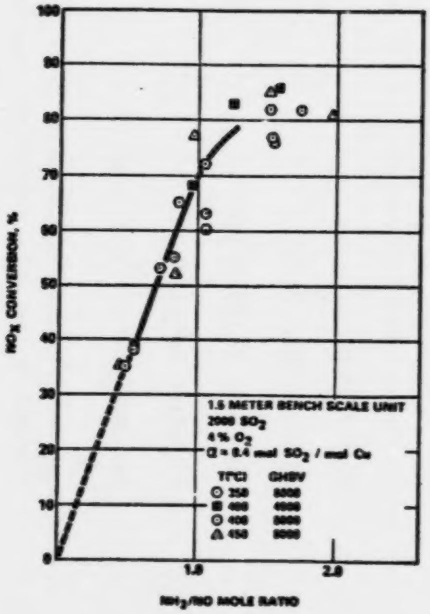


Figure 2.3.5-5. $NO_x$ reduction with $NH_3$ over commercial SFGT acceptor.[46]

TABLE 2.3.5-1. DESIGN AND OPERATING VARIABLES FOR SFGT SYSTEM[48]

| Variable | Typical Range |
|---|---|
| Space Velocity | 5,000 - 8,000 $hr^{-1}$* |
| $NH_3$:$NO_x$ Mole Ratio | 1.0:1.0 to 1.2:1.0* |
| Flue Gas Temperature | 400°C |
| Pressure Drop | 5-6 in. $H_2O$* |
| Maximum Particulate Loading | ≥23 g/Nm |

*Actual value will depend on required removal level.

Tables 2.3.5-2 and 2.3.5-3 present the test and commercial applications of the SFGT process. The development history of the process can also be seen in these tables. In the U.S., from 1974 to 1976 a pilot scale unit at Tampa Electric Company (TECO) was operated using flue gas from a coal-fired boiler. Testing was for $SO_2$ removal only, $NO_x$ control was not attempted during this period. The process developer is currently modifying the TECO pilot unit to accommodate 7 meters of bed height, up from the previous maximum of 5 meters. This should permit simultaneous removal of $NO_x$ and $SO_x$ to the 90 percent level. Also, provisions are being made for injection of a $CO/CO_2$ gas mixture into the regeneration gas in order to simulate medium-Btu gas from a coal gasifier.

The costs for an industrial size boiler have not been estimated. The only detailed cost estimates currently available are for coal-fired utility boilers. These are shown in Tables 2.3.5-4 through 2.3.5-6. Also shown are the estimated energy and raw material requirements.

TABLE 2.3.5-2. SFGT PROCESS, PILOT AND DEMONSTRATION UNITS

| Location/ Company | Designed By | Fuel/ Application | Size, $Nm^3/hr$ | Type of Operation | Dates | Comments |
|---|---|---|---|---|---|---|
| Shell Ref. at Pernis | Shell | Residual Fuel Oil-Proc. Heater | 600-1000 | $SO_x$-only | 1967-1972 | $SO_x$ reduction - approx. 90% |
| Rotterdam Utility | Shell | Coal-Steam Boiler | - | Heavy Fly Ash Loading | 1971 | Particulate matter - loadings to 20 $Gr/Nm^3$ |
| Tampa Elec. Big Bend | UOP | Coal-Wet-Bottom Utility Boiler | 1200-2000 | $SO_x$-only;<br>$SO_x$-$NO_x$ Simultaneous | 1974-1976<br>1979- | $SO_x$ - 90%;<br>$SO_x$-$NO_x$ - 90/90% fly ash to 25 $Gr/Nm^3$ |
| JGC Yokohama | JGC* | Fuel Oil | 250-700 | $NO_x$-only | 1974- | $NO_x$ reduction - 90-99% |
| Nippon Steel | JGC | Sintering Furnace | 2000 | $NO_x$-only | 1976-1978 | $NO_x$ reduction - 90-97% |
| -- | JGC | Coke Oven Gas | 400 | $NO_x$-only | 1976-1977 | $NO_x$ reduction - 90%; special low temp. cat. evaluation |

*JGC Corporation, licensing agent in Japan.

TABLE 2.3.5-3. SFGT PROCESS, COMMERCIAL UNITS

| Unit | Designed By | Fuel/ Application | Size, $Nm^3/hr$ | Type of Operation | Dates | Comments |
|---|---|---|---|---|---|---|
| SYS* Yokkaichi | Shell | Residual Fuel Oil-Ref. Boiler | 125,000 | $SO_x$-only;<br>$NO_x$-$SO_x$ Simultaneous | 1973-1975<br>1975- | $SO_x$ reduction - 90%;<br>Simultaneous - 90/50% |
| Kashima Oil Co. Ltd. | JGC | Fuel Oil-Process Unit Heater | 50,000 | $NO_x$-only | 1975- | 95-98% |
| Fuji Oil Co. Ltd. | JGC | CO Boiler | 70,000 | $NO_x$-only | 1976- | 93-96% |
| Nippon Steel Corp. | JGC | Sintering Furnace | 150,000 | $NO_x$-only | 1978- | $\sim$95% (low temp. catalyst) |

*Showa Yokkaichi Sekiyu

## TABLE 2.3.5-4. ECONOMICS OF SFGT SYSTEM[49]

| | |
|---|---|
| | **BASIS:** |
| Incorporated Units: | Steam-Naphtha Reformer<br>SFGD Reactor Section<br>Compressor/Gasholder Flow Smooth Section<br>Modified Claus Unit |
| Power Plant Size | 500 MW |
| Fuel | Coal |
| S-Content, Wt-% | |
| Case 1 | 3.5 |
| Case 2 | 2.5 |
| Case 3 | 0.8 |
| HHV | 10,500 Btu/lb |
| Heat Rate | 9,000 Btu/kWh |
| Excess Air | 20% |
| Air Preheater Leakage | 13% |
| Flue Gas Rate | 1,582,000 Nm$^3$/h (983,000 SCFM) |
| $SO_2$ Content, ppmv | |
| Case 1 | 2,580 |
| Case 2 | 1,850 |
| Case 3 | 590 |
| Mid-1977, Gulf Coast Location | |
| Load Factor | 7,000 h/a |
| Capital Charges | 15%/a |
| Cost of: | |
| Naphtha | \$0.35/gal |
| Steam (40 psi, SAT.) | \$1.50/M lb |
| Electricity | \$0.018/kWh |
| Labor | \$10.00/hr |
| Heat Credits | \$2.50/MMBtu |
| Sulfur | \$45.00/ton |

## TABLE 2.3.5-5. ECONOMICS OF SFGT SYSTEM ESTIMATED CHEMICALS AND UTILITY REQUIREMENTS[50]

| | | SFGD Section | Flow Smooth Section | Mod. Claus Section | Reformer Section | Total |
|---|---|---|---|---|---|---|
| **Case 1** | | | | | | |
| Electricity | kW | 5,770 | 850 | 115 | 480 | 7215 |
| Steam** | kmol/h | 1,820 | -380* | -740* | -600* | 100 |
| Naphtha*** | Gcal/h | | | | 90.92 | 90.92 |
| Heat Credits | Gcal/h | | | | | 42.53 |
| S° Produced | kg/h | | | 5250 | | 5250 |
| **Case 2** | | | | | | |
| Electricity | kW | 5,800 | 570 | 82 | 300 | 6782 |
| Steam** | kmol/h | 1,300 | -270* | -530* | -415* | 85 |
| Naphtha*** | Gcal/h | | | | 62.75 | 62.75 |
| Heat Credits | Gcal/h | | | | | 32.48 |
| S° Produced | kg/h | | | 3760 | | 3760 |
| **Case 3** | | | | | | |
| Electricity | kW | 5,120 | 180 | 30 | 110 | 5440 |
| Steam** | kmol/h | 480 | -95* | -170* | -140* | 75 |
| Naphtha*** | Gcal/h | | | | 21.01 | 21.01 |
| Heat Credits | Gcal/h | | | | | 18.46 |
| S° Produced | kg/h | | | 1200 | | 1200 |

*Produced

**40 psig, Saturated

***5.175 MMBtu/Bbl produces 11,500 SCF Hydrogen/Bbl

TABLE 2.3.5-6. ECONOMICS OF SFGT SYSTEM ESTIMATED CAPITAL AND OPERATING COST[51]

| | Case 1 | Case 2 | Case 3 |
|---|---|---|---|
| EEC. (MM$) | | | |
| SFGD Reactor Section | 28.95 | 28.53 | 22.94 |
| Compressor/Gasholder | 7.82 | 6.10 | 2.65 |
| Modified Claus | 2.76 | 2.26 | 1.14 |
| Steam-Naphtha Reformer | 8.81 | 7.14 | 4.17 |
| Estimated Annual Revenue Requirements (M$/a) | | | |
| Capital Charges | 7251 | 6604 | 4634 |
| Maintenance | 967 | 881 | 618 |
| Labor | 123 | 123 | 123 |
| Acceptor | 1479 | 1053 | 411 |
| Electricity | 909 | 855 | 685 |
| Steam | 42 | 35 | 31 |
| Naphtha | 7174 | 4951 | 1658 |
| Heat Credits | -2977 | -2273 | -1292 |
| Sulfur Credits | -1570 | -1126 | -359 |
| Capital Cost, Operating Cost, Energy Requirement | | | |
| Capital Cost, $/kW | 97 | 88 | 62 |
| Operating Cost, ¢/kWh | 0.38 | 0.32 | 0.19 |
| Energy Requirement, Btu/kWh* | 525 | 371 | 124 |

*Defined as the sum of:

| | |
|---|---|
| Electricity at | 9000 Btu/kWh |
| Steam at | 40000 Btu/kmol |
| Naphtha at | 4 Btu/kcal |
| Heat Credits at | 4 Btu/kcal |

#### 2.3.5.2 System Performance--

$NO_x$ control by the SFGT process is shown graphically in Figure 2.3.5-5. and in Figure 2.3.5-3 presented earlier. As can be seen, at a space velocity of 8000 $hr^{-1}$, $NO_x$ control of >80 percent can be achieved. Figure 2.3.5-4 indicates that the process developers feel the process to be capable of $NO_x$ control levels of >90 percent. Ando indicates that $NO_x$ and $SO_x$ removals of 70 percent and 90 percent, respectively are achievable at an $NH_3/NO$ mole ratio of 0.99.[120] He also indicates that higher $NO_x$ control may be possible. But unless some process modifications are made, $SO_2$ control will decrease and $NH_3$ emissions will increase.

As mentioned earlier, the system is not impacted by changes in the boiler gas rate or particulate concentrations. Changes in the $NO_x$ concentration due to boiler load changes and be compensated for by conventional control system used in conjunction with the $NH_3$ injection equipment. This control system will be developed during the upcoming pilot tests at the TECO pilot plant.

### 2.3.6 Adsorption

#### 2.3.6.1 System Description--

The adsorption process removes $NO_x$ and $SO_2$ from flue gas by adsorbing them onto a special activated char. Adsorbed $NO_x$ is reduced to $N_2$ while $SO_2$ is reduced and condensed to an elemental sulfur by-product.

A process flow diagram is shown in Figure 2.3.6-1. Flue gas is taken from the boiler air preheater and passed through a particulate removal device to prevent blinding of the adsorption bed. The flue gas then enters the adsorber, a vertical column with parallel louver beds containing the char in pellet form. $NO_x$ and $SO_2$ are adsorbed on the char which slowly moves downward through the bed. The $NO_x$ adsorption mechanism is unknown but $SO_2$ undergoes the following reaction.[55]

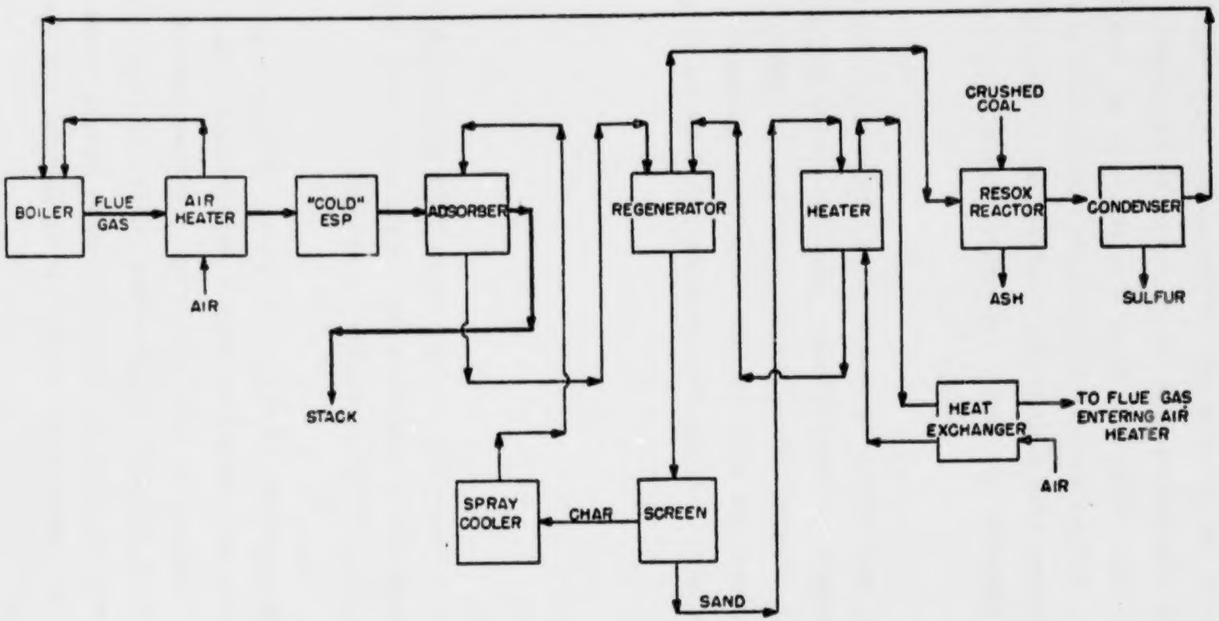

BOILER
FLUE GAS
AIR HEATER
AIR
"COLD" ESP
ADSORBER
STACK
REGENERATOR
HEATER
CRUSHED COAL
RESOX REACTOR
ASH
CONDENSER
SULFUR
HEAT EXCHANGER
TO FLUE GAS ENTERING AIR HEATER
AIR
SPRAY COOLER
CHAR
SCREEN
SAND

Figure 2.3.6-1. Flow diagram of Foster Wheeler-Bergbau Forschung Dry Adsorption Process.[54]

$$SO_2(g) + H_2O(g) + \tfrac{1}{2}O_2(g) \rightarrow H_2SO_4(l) \qquad (2\text{-}18)$$

The reaction product is held in the pores of the char pellets. The flue gas exits the adsorber and passes to the stack. The saturated char leaves the bottom of the adsorber and is screened to remove any fly ash deposits. It is then conveyed to a regenerator where it is mixed with hot sand (650°C) and the following reactions take place.[55,56]

$$2H_2SO_4(l) + C(s) \rightarrow CO_2(g) + 2H_2O(g) + 2SO_2(g) \qquad (2\text{-}19)$$

$$2NO(g) + C(s) \rightarrow CO_2(g) + N_2(g) \qquad (2\text{-}20)$$

This $SO_2$-rich gas product stream is sent to an off-gas treatment reactor containing hot, crushed coal (650-820°C) and the following reactions take place.[56]

$$SO_2(g) \rightarrow S(g) + O_2(g) \qquad (2\text{-}21)$$

$$C(s) + O_2(g) \rightarrow CO_2(g) \qquad (2\text{-}22)$$

The gas then passes to a condenser where the S vapor forms molten S. The char/sand mixture from the regenerator is screened to separate the two solids. The char is recycled to the adsorber via a spray cooler and the sand is recycled to the regenerator after passing through a heater.

This process operates at 120-150°C, however, typical values for other operating variables were not found. $NO_x$ and $SO_2$ control levels were reported to be 40-60 percent and 80-95 percent, respectively.[57] The economics of the process vary with the fuel sulfur level. For coal-fired boilers with fuel sulfur levels of 0.9-4.3 percent, the capital costs range from $40-90/kW and the operating costs range from 1.0-2.3 mills/kWh.[58] The costs were based on applying the process to a utility boiler of >200 MW capacity. Costs for oil-fired applications were not found.

Presently, the adsorption process is in the prototype unit stage of development. The one reported process developer in the field, Foster Wheeler-Bergbau Forschung has a 20 MW prototype unit and several small pilot plants treating coal-fired flue gas. The process should also be applicable to oil-fired boilers.

#### 2.3.6.2 System Performance--

Tests have shown the adsorption process to be primarily a $SO_2$ reduction process as $NO_x$ removal efficiency averages 40-60 percent while $SO_2$ removal had a range of 80-95 percent.[59]

The primary drawback of this process, besides the low $NO_x$ removal level, is its complexity: numerous process steps involving hot solids handling. Solids flow can be difficult to control and high maintenance requirements could be expected. The vendor has reported several mechanical problems during testing which included control of adsorber-bed levels, poor char distribution, char-sand separation, hot sand conveying, and char cooling and feed. Some corrosion-resistant material is needed in the high temperature zones of the process. The ash waste stream from the off-gas treatment reactor appears to be the sole secondary pollutant associated with the process. The overall complexity and low $NO_x$ removal of the process present definite technical disadvantages.

### 2.3.7 Electron Beam Radiation

#### 2.3.7.1 System Description--

This dry process utilizes an electron beam to bombard the flue gas, removing $NO_x$ and $SO_2$ in the process. A block flow diagram for the process is shown in Figure 2.3.7-1.

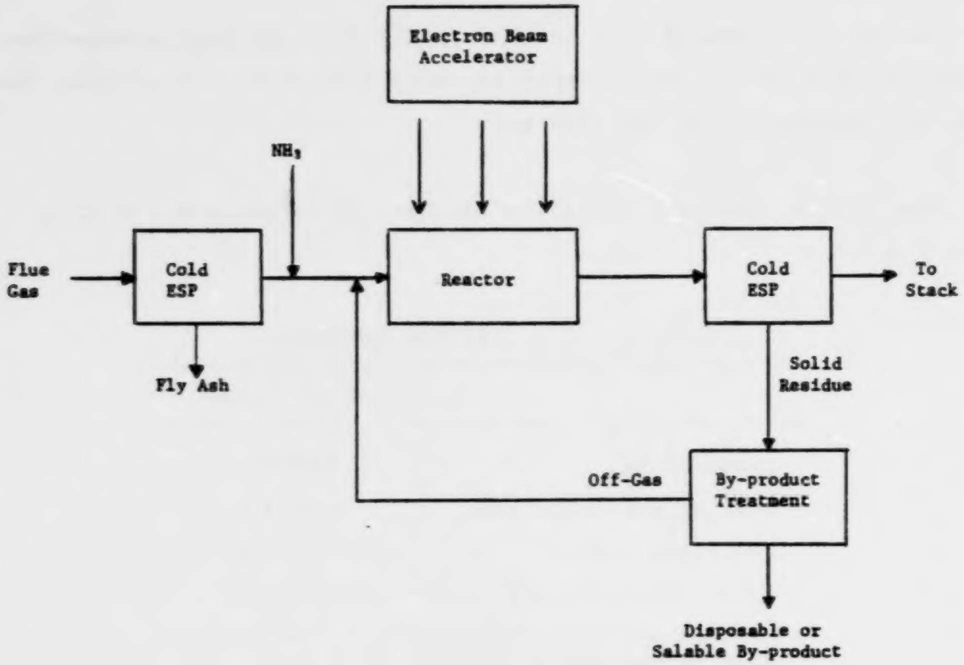

Electron Beam Accelerator
NH$_3$
Flue Gas
Cold ESP
Reactor
Cold ESP
To Stack
Fly Ash
Solid Residue
Off-Gas
By-product Treatment
Disposable or Salable By-product

Figure 2.3.7-1. Process flow diagram for Ebara-JAERI electron beam process.[60]

Flue gas is taken from the boiler air preheater and passed through a cold ESP to remove particulates. After a small amount of ammonia is added, the gas enters a reactor where it is bombarded with an electron beam. (The penetration of the gas stream by the beam will require a unique discharge pattern or other special design considerations.) A powder containing both ammonium nitrate and sulfate is generated by an unknown reaction mechanism. The gas then exits the reactor, passes through a second ESP to remove the solid by-product, and is sent to the stack. The by-product treatment system is still being developed. Various methods investigated include thermal decomposition in the presence of an inert gas, steam roasting with CaO, or steam roasting with $H_2O$. The by-product may eventually be useful as a fertilizer.[61]

The key subsystem of this process is the electron beam accelerator. Control of this unit's power supply is based upon inlet composition, flow rate, and temperature of the flue gas.

Some of the important variables and typical ranges are listed in Table 2.3.7-1.

TABLE 2.3.7-1. SYSTEM VARIABLES[62]

| | Typical Value |
|---|---|
| Temperature | ∿100°C |
| Reactor residence time | 1-20 sec |
| Radiation rate | $10^5$-$10^6$ rad*/sec |
| Total radiation absorbed | 1-3 Mrad* |

*Rad is the radiation dose absorbed
1 rad = .01 J/Kg

The operating cost with $NO_x$ removal only (low sulfur oils) is lower due to lower radiation levels, but the capital cost would be just as high as for simultaneous $NO_x/SO_x$ removal. Capital costs are quite high for this process as the 2 ESP's and the accelerator are expensive. The costs for a 1000 $Nm^3$/hr test unit are reported to be $1000/kW; however, the costs of a full scale system are expected to be lower. Operating costs are not available.

The Ebara Manufacturing Company in conjunction with Japan Atomic Energy Research Institute (JAERI) has operated a 1000 $Nm^3$/hr pilot plant treating flue gas from an oil-fired boiler. In 1976, a 3000 $Nm^3$/hr pilot plant began treating off-gas from an iron ore sintering furnace at Nippon Steel. By-product treatment technology needs to be more fully developed before this process can be applied commercially.

In the U.S., the Department of Energy (DOE) is funding development of an electron beam process offered by Research-Cottrell. Pilot unit tests with flue gas are scheduled, however, the details of the program are not yet available.

### 2.3.7.2 System Performance--

A summary of the oil-fired pilot tests is shown in Figure 2.3.7-2.

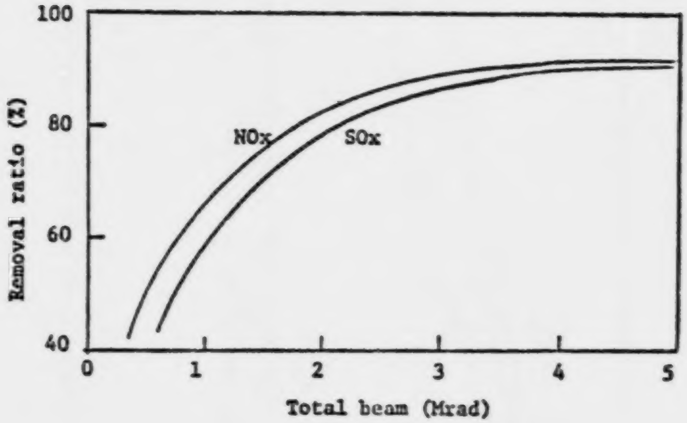


Figure 2.3.7-2. Oil-fired pilot plant test results.[64]

One can see that $NO_x/SO_x$ removal drops off drastically at a total radiation dose below 1 Mrad while the maximum removal is obtained at about 3 Mrad. The removal efficiencies decrease as the concentrations of $NO_x$ and $SO_x$ increase as can be seen in Figure 2.3.7-3.

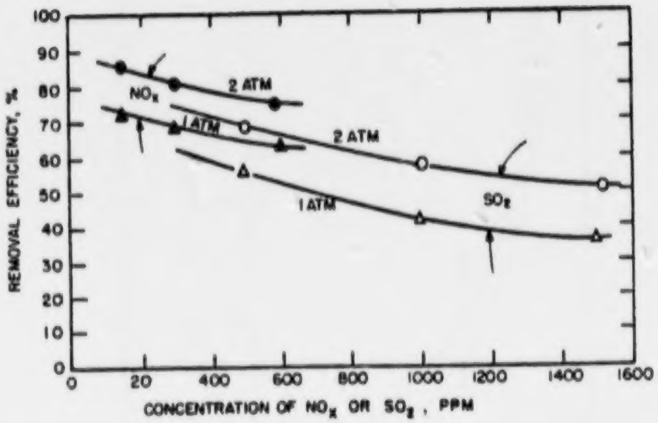


Figure 2.3.7-3. Effect of pollutant concentration on removal efficiency.[65]

## 2.3.8 Absorption-Reduction

### 2.3.8.1 System Description--

Absorption-reduction processes simultaneously remove $NO_x$ and $SO_2$ from flue gas by absorbing them into a scrubbing solution. The processes are based on the use of chelating compounds, such as ethylenediamine tetraacetic acid (EDTA) complexed with iron, to "catalyze" the absorption of $NO_x$. Most process vendors prefer a perforated-plate type of gas-liquid contactor. The advantages of a perforated-plate absorber over a packed bed absorber include easier cleaning when solids are present, wider operating ranges, and more economical handling of high liquid rates.[66] An example of a perforated plate absorber is shown in Figure 2.3.8-1. The most common design of a perforated plate is one that employs liquid crossflow over the face of the plate with the gas passing upward through the plate perforations. A schematic of the operation of a crossflow perforated plate is shown in Figure 2.3.8-2. The liquid is prevented from flowing through the plates by the upward flow of the gas. However, during periods of low gas flow (such as load changes on industrial boilers) liquid can drain through the openings in the plates. This reduces the liquid's time of contact with the gas on each plate and may decrease the overall operating efficiency of the absorber. To prevent this problem, there are two other types of dispersers utilized besides the basic sieve-plate: the valve-plate and the bubble cap, depicted in Figure 2.3.8-3. As the gas flow lowers, the valve or cap settles, sealing off the perforation so liquid cannot drain through. This design feature allows the perforated plate absorber to operate more efficiently at widely fluctuating gas rates.

While most all absorption-reduction processes utilize ferrous chelating compounds to enhance NO absorption, the scrubbing solutions, the by-product treatment and sorbent regeneration chemistry differ from process to process. For this reason, one of the simpler absorption-reduction processes, that of Kureha Chemical Industry Company, is examined here in detail.

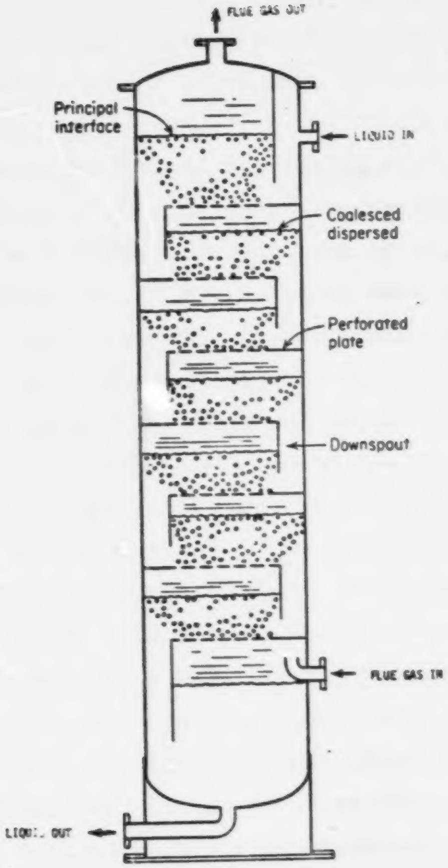

FLUE GAS OUT
Principal interface
LIQUID IN
Coalesced dispersed
Perforated plate
Downspout
FLUE GAS IN
LIQUID OUT

Figure 2.3.8-1. Perforated plate absorber option for Absorption-Reduction Processes.[28]

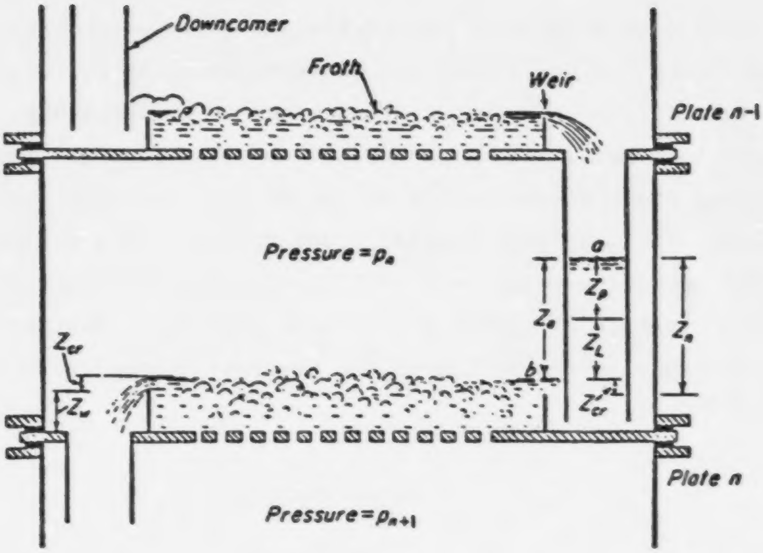
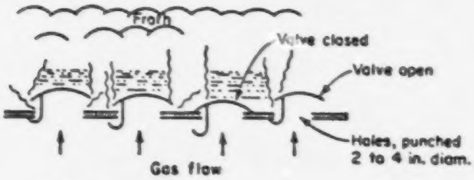
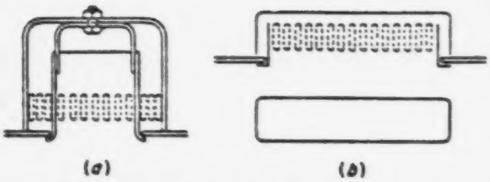

Downcomer
Froth
Weir
Plate n-1
Pressure = $p_a$
a
$Z_p$
$Z_a$
$Z_L$
$Z_a$
b
$Z_{cr}$
$Z_{cr}$
$Z_w$
Plate n
Pressure = $p_{n+1}$

Figure 2.3.8-2. Normal operation of sieve plate. $Z_a$, height of station a above datum. $Z_{cr}$, weir crest. $Z_L$, liquid-friction head. $Z_p$, pressure head across plate. $Z_n$, net head in down pipe. $Z_w$, weir height.[67]


Froth
Valve closed
Valve open
Holes, punched
2 to 4 in. diam.
Gas flow

Valve-plate dispersers.



(a) Circular or bell cap. (b) Tunnel cap.

Bubble cap dispersers

Figure 2.3.8-3. Other gas dispersers.[68]

A block flow diagram of the Kureha absorption-reduction process is shown in Figure 2.3.8-4. Flue gas is taken from the boiler after the air preheater. It passes through a prescrubber to adiabatically cool the gas and remove both particulates and chlorides. The flue gas then enters the distributing space at the bottom of the $NO_x/SO_2$ absorber, below the plates or packing. The gas flows upward, countercurrent to a sodium acetate ($CH_3COONa$) scrubbing solution (∿60°C) containing ferrous iron and EDTA and a few seed crystals of gypsum (to prevent scaling). Most of the $SO_2$ is rapidly absorbed at the bottom of the absorber according to the following reactions.[71]

$$SO_2(g) \rightarrow SO_2(aq) \qquad (2\text{-}23)$$

$$SO_2(aq) + 2CH_3COONa(aq) + H_2O \rightarrow Na_2SO_3(aq) + 2CH_3COOH(aq) \qquad (2\text{-}24)$$

The $NO_x$ (which consists mainly of NO) is relatively insoluble; therefore, it is absorbed gradually over the length of the column. The ferrous chelating compounds effect on NO absorption is described in Figure 2.3.8-5. The $NO_x$ is absorbed and undergoes the following reactions.[73]

$$NO(g) \rightarrow NO(aq) \qquad (2\text{-}6)$$

$$2NO_2(g) \rightarrow N_2O_4(aq) \qquad (2\text{-}25)$$

$$N_2O_4(g) \rightarrow N_2O_4(aq) \qquad (2\text{-}9)$$

$$2NO(aq) + 5Na_2SO_3(aq) + 4CH_3COOH(aq) \rightarrow 2NH(SO_3Na)_2(aq) + Na_2SO_4(aq) + 4CH_3COONa(aq) + H_2O \qquad (2\text{-}26)$$

$$2N_2O_4(aq) + 7Na_2SO_3(aq) + 4CH_3COOH(aq) \rightarrow 2NH(SO_3Na)_2(aq) + 3Na_2SO_4(aq) + 4CH_3COONa(aq) + H_2O \qquad (2\text{-}27)$$

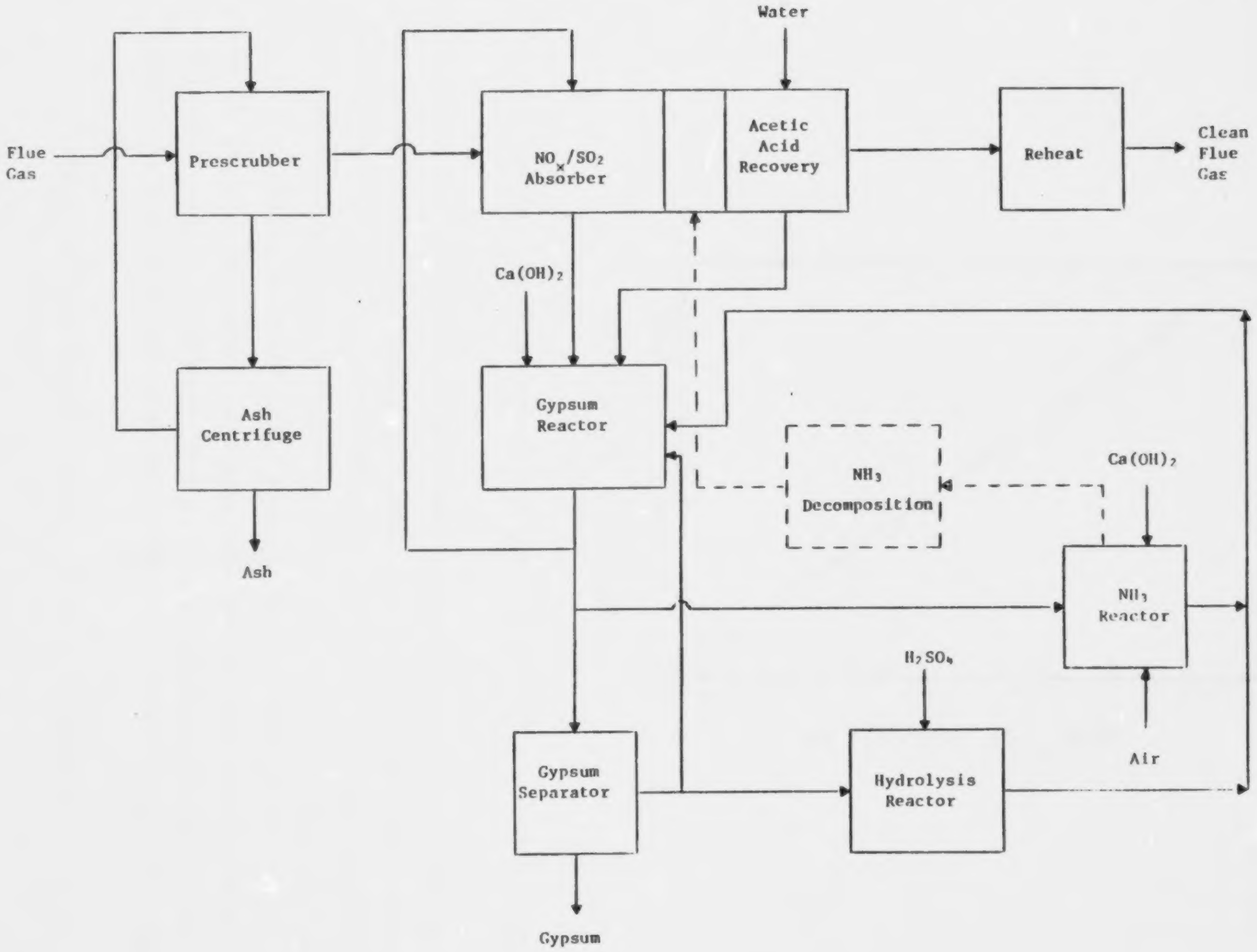

Flue Gas
Prescrubber
Ash Centrifuge
Ash
NO$_x$/SO$_2$ Absorber
Acetic Acid Recovery
Water
Reheat
Clean Flue Gas
Ca(OH)$_2$
Gypsum Reactor
NH$_3$ Decomposition
Ca(OH)$_2$
NH$_3$ Reactor
Air
H$_2$SO$_4$
Gypsum Separator
Hydrolysis Reactor
Gypsum

Figure 2.3.8-4. Process flow diagram of Kureha absorption-reduction process.[69,70]

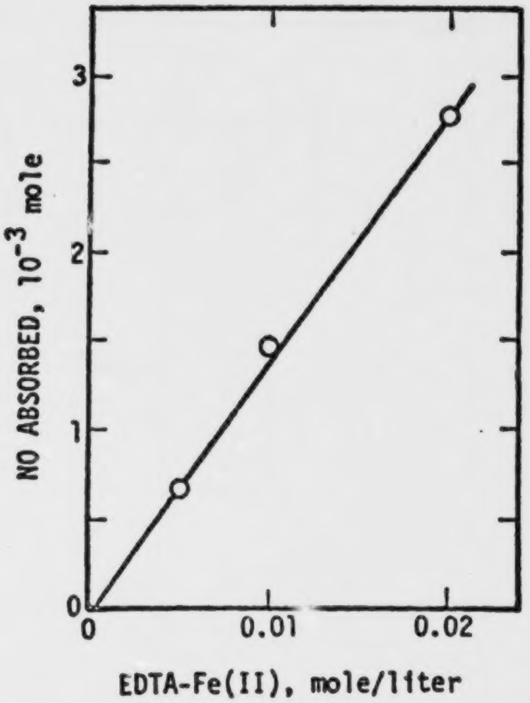


Figure 2.3.8-5. EDTA-Fe(II) concentration and NO absorption at 50°C.[72]

Some of the acetic acid ($CH_3COOH$) formed at the bottom of the absorber via reaction (2-24) is vaporized. It must be captured and is done so by water scrubbing at the very top of the absorber. From the top of the absorber column the clean flue gas passes to a heater for plume buoyancy and is then sent to the stack. The liquid effluent drops from the bottom of the absorber to a gypsum, $CaSO_4 \cdot 2H_2O$, production reactor. Here, the solution is mixed with with the purge stream from the acetic acid recovery section and a lime slurry stream. The lime, $Ca(OH)_2$, treatment involves the following reactions.[74]

$$2CH_3COOH(aq) + Ca(OH)_2(aq) \rightarrow (CH_3COO)_2Ca(aq) + 2H_2O \qquad (2\text{-}28)$$

$$(CH_3COO)_2Ca(aq) + Na_2SO_4(aq) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \downarrow + 2CH_3COONa(aq) \quad (2\text{-}29)$$

The gypsum formed by reaction 2-29 is centrifuged. Most of the liquor discharged is returned to the gypsum reactor and on to the absorber. The remaining liquor is sent to a reactor where sulfuric acid ($H_2SO_4$) is added to hydrolyze the imidodisulfonate, $NH(SO_3Na)_2$, by the following reaction.[75]

$$NH(SO_3Na)_2(aq) + 2H_2O \xrightarrow[140°C]{H^+} NH_4HSO_4(aq) + Na_2SO_4(aq) \qquad (2\text{-}30)$$

The effluent from this reactor is then recycled to the gypsum production reactor. A small purge stream is taken from the gypsum reactor to another reactor where the ammonium bisulfate ($NH_4HSO_4$) formed in the hydrolysis reaction is treated with lime to yield gypsum and $NH_3$ off-gas by the following reaction.[76]

$$NH_4HSO_4(aq) + Ca(OH)_2(s) \rightarrow CaSO_4 \cdot 2H_2O(s)\downarrow + NH_3(g)\uparrow \qquad (2\text{-}31)$$

The gaseous ammonia is stripped from the solution by an air stream. If no use for the ammonia can be found, the gas mixture is sent to a catalytic reactor where ammonia is oxidized by the following reaction.[77]

$$4NH_3(g) + 3O_2(g) \xrightarrow[350°C]{catalyst} 2N_2(g) + 6H_2O(g) \qquad (2\text{-}32)$$

The product stream is then sent to the deacetating section of the absorber column.

The fundamental design equation used for gas absorption column design is[32]

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \qquad (2\text{-}11)$$

where $y$ = bulk $NO_x$ concentration (mole fraction of gas phase at any given point in column

$y-y^*$ = overall driving force for absorption ($y^*$ being the $NO_x$ concentration of a gas in equilibrium with given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/($ft^2$)(hr)(mole fraction)

$a$ = area of gas-liquid interface per unit packed volume, $ft^2/ft^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/($ft^2$)(hr)

$Z$ = length of packed section of column, ft

In a column containing a given plate or packing configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity, obviously, must be lower than the flooding velocity. How much lower is a design decision. Often it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[33] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter consideration is the reason why a chelating compound is used in absorption-reduction processes to aid in $NO_x$ absorption. Table 2.3.8-1 presents the effects of boiler/flue gas variables on the design of absorption-reduction systems. Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load change. The column size and the liquid flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

TABLE 2.3.8-1. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Requires $SO_2$:$NO_x$ mole ratio of at least 3-5[9] (depending on process) for absorption-reduction to be effective. |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased catalyst concentration |

The process vendors have not released much information on the operating conditions of these processes. This is primarily due to the competitive status of these similar processes at this early stage of development. Typical values for some of the process variables are shown in Table 2.3.8-2.

TABLE 2.3.8-2. TYPICAL VALUES FOR PROCESS VARIABLES OF ABSORPTION-REDUCTION PROCESSES[78]

| Variable | Range |
|---|---|
| Liquid/Gas ratio, $1/Nm^3$ | 10-30 |
| $SO_x/NO_x$ mole ratio | 2.5-3.0 |
| Superficial Gas Velocity, m/s | 1-3 |

Cost estimates for this type of process cover a large range, presumably due to the differences in sorbent regeneration technique. Capital costs for an oil-fired system were not found, however, for coal-fired utility applications, capital costs ranged from \$65-127/kW. Operating costs for an oil-fired application are shown below for two flue gas flow rates.[121]

| Gas Flow Rate, $Nm^3$/hr | Operating Costs, mills/kWh |
|---|---|
| 150,000 | 9.1 |
| 500,000 | 8.3 |

These costs are based on ¥200/\$ and do not include flue gas reheat.

Presently, absorption-reduction processes are in the pilot-unit stage of development. Table 2.3.8-3 presents a list of absorption-reduction process vendors and the status of development of their projects. One can see from the table that several oil-fired flue gas tests have been performed.

TABLE 2.3.8-3. PROCESS VENDORS OF ABSORPTION-REDUCTION PROCESSES[80]

| Vendor | Status of Development |
|---|---|
| Asahi | 1974: 600 $Nm^3$/hr flue gas from residual oil-fired boiler (1000 hours continuous). |
| Chisso | 1975: 300 $Nm^3$/hr flue gas from oil-fired boiler (335 hours continuous) |
| Kureba | 1976: 5000 $Nm^3$/hr flue gas from heavy oil-fired boiler (3000 hours continuous) |
| Mitsui Engineering and Shipbuilding | 1974: 150 $Nm^3$/hr flue gas from oil-fired boiler |
| Pittsburgh Environmental | 1976: 3000 $Nm^3$/hr flue gas from coal-fired boiler (52 hours continuous, absorption section only) |

### 2.3.8.2 System Performance--

Four of the vendors listed in Table 2.3.8-3 report $NO_x$ removals of at least 80 percent with oil-fired boiler flue gas. The Pittsburgh Environmental system, however, only achieves 60 percent with coal. $SO_2$ control for all of the systems is 90+ percent. No plots of system performance could be found for any of the oil-fired systems.

Absorption-reduction processes are readily applicable only to high sulfur oils as a $SO_2:NO_x$ mole ratio in the flue gas of at least 3-5 is required for maximum performance. This can easily be shown by observing reactions 2-24 and 2-26 reprinted below.

$$SO_2(aq) + 2CH_3COONa(aq) + H_2O \rightarrow Na_2SO_3(aq) + 2CH_3COOH(aq)$$

$$\begin{aligned}2NO(aq) + 5Na_2SO_3(aq) + 4CH_3COOH(aq) \rightarrow 2NH(SO_3Na)_2(aq) &+ Na_2SO_4(aq) \\ &+ 4CH_3COONa(aq) + H_2O\end{aligned}$$

One can see that 1 mole of $SO_2$ absorbed in solution reacts to form 1 mole of sodium sulfite ($Na_2SO_3$). Then, 5 moles of sodium sulfite are required to reduce 2 moles of NO. So, the minimum stoichiometric $SO_2:NO_x$ mole ratio required is $\frac{5}{2}$ or 2.5. Also, some of the sodium sulfite is oxidized to sodium sulfate by oxygen present in the flue gas according to:

$$Na_2SO_3(aq) + \tfrac{1}{2}O_2(aq) \rightarrow Na_2SO_4(aq) \qquad (2\text{-}33)$$

and is not available for $NO_x$ reduction. Low-sulfur oils would require $SO_2$ to be added to the flue gas for these processes to perform; therefore, they should be considered applicable to high sulfur oils only.

Absorption-reduction processes require large absorbers with high liquid rates due to relative insolubility of NO, even when the absorption catalyst is used. Also, the regeneration of the absorption catalyst and the flue gas reheat for plume buoyancy are energy intensive. Some corrosion-resistant material is necessary due to the corrosive nature of the absorbing solution. However, absorption-reduction appears to be the most promising of the "wet" $NO_x/SO_2$ removal processes. This is due primarily to its not utilizing oxidants which require much corrosion-resistant material and, more importantly, create serious secondary pollution problems. Also, the primary by-product of absorption-reduction processes, gypsum, can be used as landfill material or in building materials. For all the above reasons, absorption-reduction processes appear, at this preliminary stage, to be competitive with other wet $NO_x/SO_x$ removal processes.

## 2.3.9 Oxidation-Absorption-Reduction

### 2.3.9.1 System Description--

Oxidation-absorption-reduction processes simultaneously remove $NO_x$ and $SO_2$ from flue gas by oxidizing relatively insoluble NO to relatively soluble $NO_2$ and then absorbing both $NO_2$ and $SO_2$ into a scrubbing solution. The processes are based on the use of gas-phase oxidants, either ozone ($O_3$) or chlorine dioxide ($ClO_2$), to selectively oxidize NO to $NO_2$. Both perforated-plate and packed bed absorption columns are utilized by various process vendors.

Most of the oxidation-absorption-reduction processes are similar in that they consist of five major sections:

- prescrubbing
- gas-phase oxidation
- $NO_x/SO_2$ absorption
- reduction of absorbed $NO_x$ and oxidation of $SO_3^=$
- wastewater treatment

The areas where processes differ are gas-phase oxidation - $O_3$ or $ClO_2$; absorption solutions - limestone slurry ($CaCO_3$), $H_2SO_4$, or NaOH; and the amount and type of waste treatment required. Thermal decomposition, biological denitrification, or wastewater evaporation wastewater treatment systems can be used. Because of these differences, only one of the oxidation-absorption-reduction processes, that of Mitsubishi Heavy Industries, is examined here in detail.

A block flow diagram of the MHI oxidation-absorption-reduction process is shown in Figure 2.3.9-1.

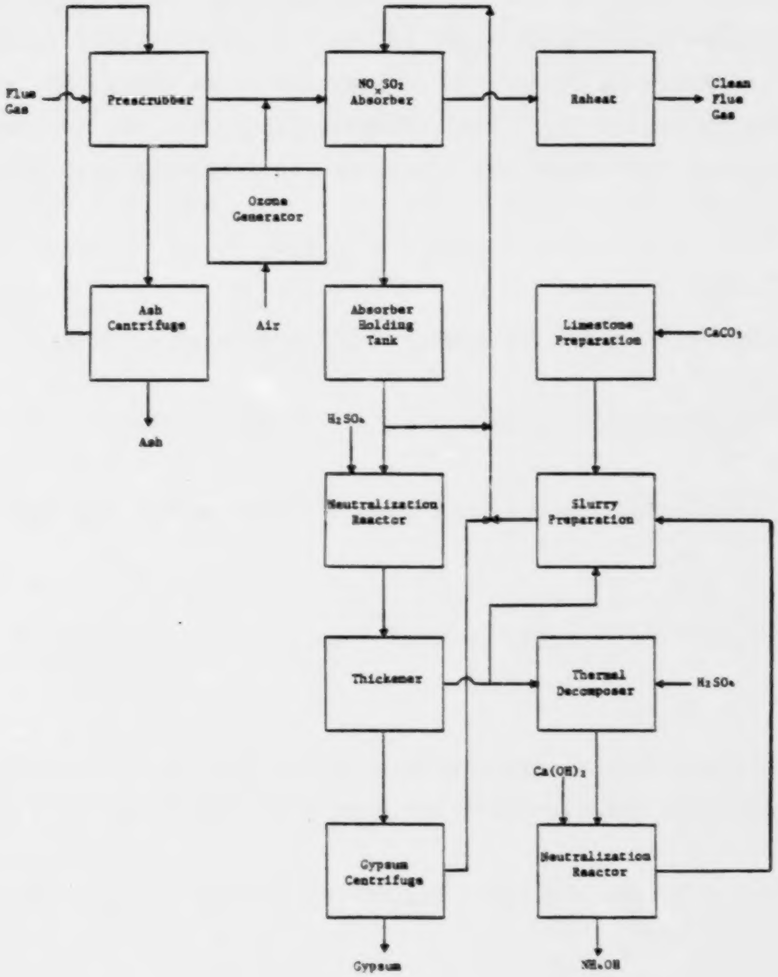

Flue Gas
Prescrubber
$NO_x$ $SO_2$ Absorber
Reheat
Clean Flue Gas
Ozone Generator
Ash Centrifuge
Air
Absorber Holding Tank
Limestone Preparation
$CaCO_3$
Ash
$H_2SO_4$
Neutralization Reactor
Slurry Preparation
Thickener
Thermal Decomposer
$H_2SO_4$
$Ca(OH)_2$
Gypsum Centrifuge
Neutralization Reactor
Gypsum
$NH_4OH$

Figure 2.3.9-1. Process flow diagram for MHI oxidation-absorption-reduction process.[81]

Flue gas is taken from the boiler after the air preheater and passed through a prescrubber to cool the gas and remove particulates and chlorides. The flue gas then enters a duct where it is injected with ozone (about 1 percent by weight in air)[82] such that the $O_3$:NO ratio is 1:1. Ozone selectively oxidizes NO by the following reatcion.[83]

$$NO(g) + O_3(g) \rightarrow NO_2(g) + O_2(g) \qquad (2\text{-}34)$$

After injection, the flue gas passes countercurrent to a lime/limestone slurry in a grid-packed absorption column. A water-soluble catalyst is added to the slurry to enhance $NO_2$ absorption (even though $NO_2$ is more soluble than NO, it is still less soluble than $SO_2$). $SO_2$ is absorbed quickly at the bottom of the column and undergoes the following reactions.[15]

$$SO_2(g) \rightarrow SO_2(aq) \qquad (2\text{-}23)$$

$$SO_2(aq) + CaCO_3(s) + \tfrac{1}{2}H_2O \rightarrow CaSO_3\cdot\tfrac{1}{2}H_2O(s) + CO_2(g) \qquad (2\text{-}35)$$

$$SO_2(aq) + CaSO_3(aq) + H_2O \rightarrow Ca(HSO_3)_2(aq) \qquad (2\text{-}36)$$

$NO_2$ is absorbed gradually over the length of the column and reacts as follows.[16]

$$2NO_2(g) + Ca(OH)_2(s) + CaSO_3\cdot\tfrac{1}{2}H_2O(s) + \tfrac{1}{2}H_2O \rightarrow Ca(NO_2)_2(aq) + CaSO_4\ 2H_2O(s) \qquad (2\text{-}37)$$

Once both the $NO_2$ and $SO_2$ are absorbed, the nitrite ion formed by reaction 2-37 is reduced by the bisulfate ion formed by reaction 2-36.[84]

$$Ca(NO_2)_2(aq) + 3Ca(HSO_3)_2(aq) \rightarrow 2Ca[NOH(SO_3)_2](aq) + 2CaSO_3\cdot\tfrac{1}{2}H_2O(s)\downarrow + H_2O \qquad (2\text{-}38)$$

These hydroxylamine $[NOH(SO_3)_2^{=}]$ compounds are reduced further by the sulfite ion[85]

$$Ca[NOH(SO_3)_2](aq) + CaSO_3\cdot\tfrac{1}{2}H_2O(s) + \frac{3}{2}H_2O \rightarrow Ca[NH(SO_3)_2](aq) + CaSO_4\cdot 2H_2O(s)\downarrow \tag{2-39}$$

Upon leaving the top of the absorber, the clean flue gas is reheated for plume buoyancy and sent to the stack. The slurry solution drops to a holding tank from which most of the solution is returned to the top of the absorber. A small stream passes to a neutralization reactor where sulfuric acid is added to convert the sulfite solid to soluble bisulfite and solid gypsum:[86]

$$2CaSO_3\cdot\tfrac{1}{2}H_2O(s) + H_2SO_4(aq) + H_2O \rightarrow CaSO_4\cdot 2H_2O(s)\downarrow + Ca(HSO_3)_2(aq) \tag{2-40}$$

This stream passes to a thickener from which the bottoms are sent to a centrifuge to separate the solid gypsum by-product from the liquor which is returned to the absorber. The overflow from the thickener is primarily recycled to the limestone slurry preparation tank. The remainder is sent to a thermal decomposer where sulfuric acid is added to hydrolyze the N-S compounds.[18]

$$2Ca[NH(SO_3)_2](aq) + 2H_2O \xrightarrow{H^+} Ca(NH_2SO_3)_2(aq) + Ca(HSO_4)_2(aq) \tag{2-41}$$

$$Ca(NH_2SO_3)_2(aq) + Ca(HSO_4)_2(aq) + 6H_2O \xrightarrow{H^+} 2NH_4HSO_4(aq) + 2CaSO_4\cdot 2H_2O(s)\downarrow \tag{2-42}$$

The ammonium bisulfate solution is pumped to another neutralization reactor where lime is added.[87]

$$NH_4HSO_4(aq) + Ca(OH)_2 + H_2O \rightarrow CaSO_4\cdot 2H_2O(s)\downarrow + NH_4OH(aq) \tag{2-43}$$

MHI has three possible methods of removing this ammonium hydroxide:

- decompose by increasing pH
- decompose thermally
- strip out with makeup $H_2O$

The remaining gypsum slurry is pumped to the limestone slurry preparation tank.

The fundamental design equation used for gas absorption column design is[32]

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \qquad (2\text{-}11)$$

where y = bulk $NO_x$ concentration (mole fraction) of gas phase at any given point in column

y-y* = overall driving force for absorption (y* being the $NO_x$ concentration of a gas in equilibrium with a given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/($ft^2$)(hr)(mole fraction)

a = area of gas-liquid interface per unit packed volume, $ft^2/ft^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/($ft^2$)(hr)

Z = length of packed section of column, ft

In a column containing a given plate or packing configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity obviously, must be lower than the flooding velocity. How much lower is a design decision. Often, it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[33] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter consideration is why oxidation-absorption-reduction processes oxidize NO to more soluble $NO_2$ prior to the absorber and why some processes add water soluble catalysts to the scrubbing solution to aid $NO_2$ absorption. The oxidation step enables these processes to use shorter absorbers with lower liquid rates than either the absorption-oxidation or absorption-reduction processes. Table 2.3.9-1 presents the effects of boiler/flue gas variables on the design of oxidation-absorption-reduction systems. Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load change. The column size and the liquid, oxidant, and catalyst flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

Typical ranges for several operating parameters for this type of process are shown in Table 2.3.9-2. Reagent concentrations were not available. Economics for the various processes cover a wide range presumably due to different techniques for oxidant generation and treatment of the scrubbing solution. Costs are reported to range from $84-134/kW for utility applications' capital expense and 6.7-9 mills/kWh for operating expense.[91]

Presently, some of the oxidation-absorption-reduction processes have reached the prototype stage of development. Table 2.3.9-3 presents a list of oxidation-absorption-reduction process vendors and the status of development of their projects. The applications of this process have been predominately to oil-fired boilers. Some of the applications treat flue gas flow rates similar to those for the standard boilers of this study. Application to industrial boilers, therefore, is technically feasible.

TABLE 2.3.9-1. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Depends on individual process: if $NO_2^-$ is completely reduced to $N_2$ or $NH_3$ by $SO_3^=$ (as does MHI), then at least the stoichiometric $SO_2:NO_x$ mole ratio of 3:1 is required[88] [see equation (9-6)]; if $NO_2^-$ is not reduced completely, then a different ratio will be necessary |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased gas-phase oxidant flow rate; increased liquid-phase catalyst concentration |

TABLE 2.3.9-2. TYPICAL RANGES OF OPERATING VARIABLES FOR OXIDATION-ABSORPTION-REDUCTION PROCESSES[89,90]

| Variable | | Range |
|---|---|---|
| Liquid/Gas Ratio, $1/Nm^3$ | | 2-12 |
| Oxidant/NO Mole Ratio | $O_3$ systems | 0.6-1.0 |
| | $ClO_2$ systems | 0.55 |
| $SO_2/NO_x$ Mole Ratio | | 2.5-5.0 |
| Superficial Gas Velocity, m/s | | 3-5 |
| Pressure Drop, $mmH_2O$ | | 200-500 |

TABLE 2.3.9-3. PROCESS VENDORS OF OXIDATION ABSORPTION-REDUCTION PROCESSES[92,93]

| Vendor | Status of Development |
|---|---|
| Chiyoda | 1975: 1000 $Nm^3$/hr flue gas from heavy oil-fired boiler |
| Ishikawajima-Harima Heavy Industries | 1975: 5000 $Nm^3$/hr flue gas from oil-fired boiler (3000 hours continuous) |
| Mitsubishi Heavy Industries | 1975: 2000 $Nm^3$/hr flue gas from heavy oil-fired boiler (700 hours continuous) |
| Osaka Soda | 1976: 60,000 $Nm^3$/hr flue gas from oil-fired boiler |
| Shirogane | 1974: 48,000 $Nm^3$/hr flue gas from oil-fired boiler |
| Sumitomo Metal-Fujikasui: Calcium Process | 1976: 25,000 $Nm^3$/hr flue gas from sintering furnace |
| Sumitomo Metal-Fujikasui: Sodium Process | 1973: 62,000 $Nm^3$/hr flue gas from heavy oil-fired boiler (5 others) |

2.3.9.2 System Performance--

Results of oil-fired tests show up to 90 percent $NO_x$ reduction and >95 percent $SO_2$ reduction. Figures 2.3.9-2 and 2.3.9-3 illustrate $NO_x$ removals as a function of pH and additive concentrations.

The primary disadvantage of these processes is the utilization of costly gas-phase oxidants which create secondary wastewater pollution problems. Both ozone and chlorine dioxide are highly unstable so they cannot be stored and must be generated onsite. $O_3$, the more expensive of the two, is generated by a high energy corona discharge in air. This instantaneous process requires significantly large amounts of electricity. $ClO_2$ is generated by a slower chemical reaction (requires about 20 minutes to respond to a change in demand) which could make it less responsive to boiler load changes. The use of $ClO_2$ introduces an additional secondary pollutant, chlorides, besides the nitrite salt problem. Significant amounts of corrosion-resistant material are required for oxidation-absorption-reduction processes, regardless of which oxidant is utilized. Some of the processes would not be applicable to low sulfur oils as they require large amounts of $SO_2$ to obtain $NO_2(aq)$ or $NO_2^-$ reduction.

CONDITIONS

CONCENTRATION OF $CaSO_3$: 5 wt%
pH OF LIQUOR : 5.5
$NO_2/(NO + NO_2)$ : 0.95

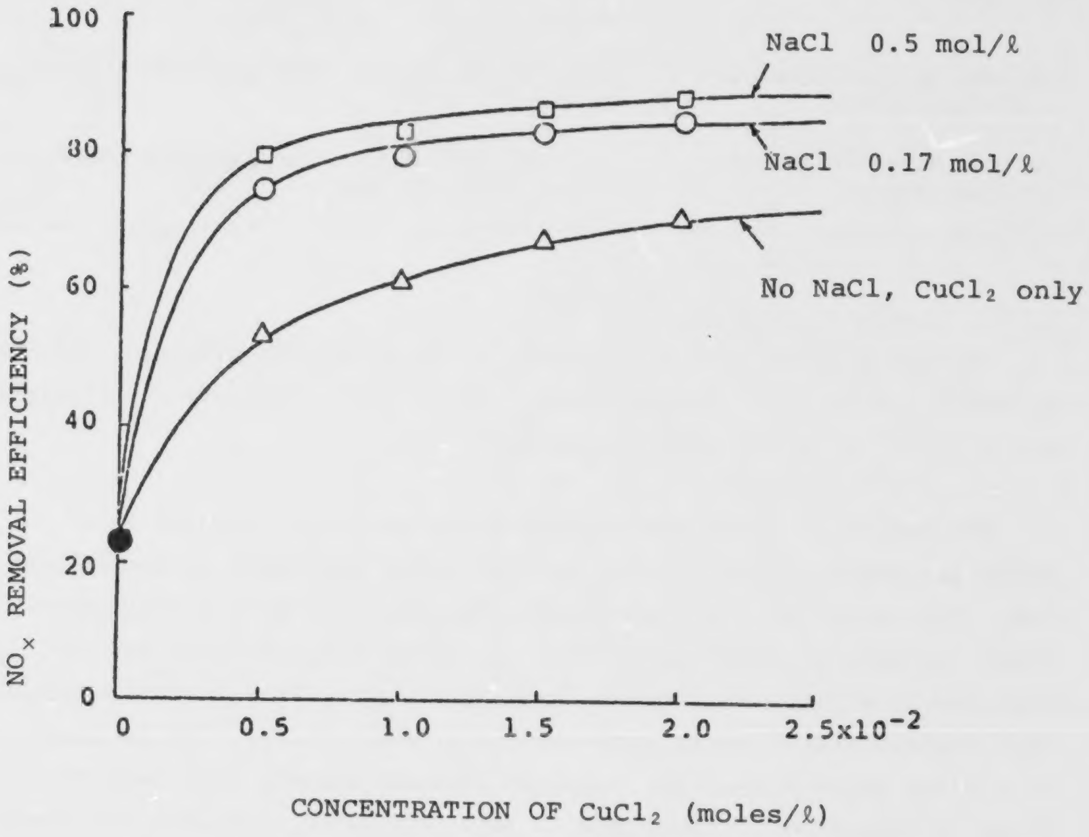


Figure 2.3.9-2. Effect of $CaCl_2$ and NaCl concentration on $NO_x$ removal efficiency.[122]

CONDITIONS

CONCENTRATION OF $CaSO_3$ : 5 wt%
CONCENTRATION OF $CuCl_2$ : 0.01 mole/ℓ
CONCENTRATION OF NaCl : 1 wt%
$NO_2/(NO + NO_2)$ : 0.95

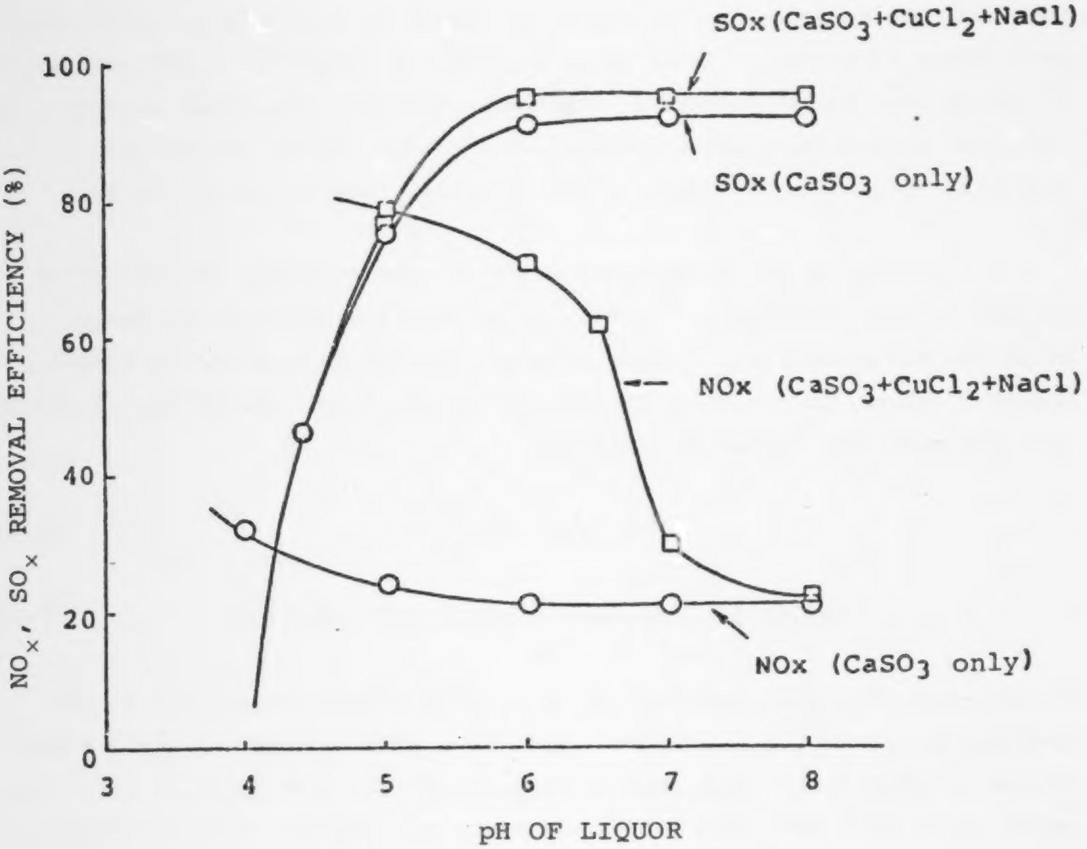


Figure 2.3.9-3. Effect of pH on $SO_x$ and $NO_x$ removal efficiency.[123]

## 2.3.10 Oxidation-Absorption

### 2.3.10.1 System Description--

As a group, oxidation-absorption processes include those oxidation processes which do not qualify for the oxidation-absorption-reduction category. Basically, there are two types of oxidation-absorption processes. One is a simplified version of the oxidation-absorption-reduction process and uses an excess of ozone to selectively oxidize $NO_x$ to $N_2O_5$ which is absorbed into aqueous solution and concentrated to form a 60 percent nitric acid ($HNO_3$) by-product. There is no reduction of $NO_x(NO_2^-)$ by the absorption of $SO_2$(as $SO_3^=$) and no wastewater treatment facility. The other type of oxidation-absorption process is based on equimolar $NO$-$NO_2$ absorption: absorbing $N_2O_3$ which is formed by the gas-phase reaction of NO and $NO_2$.

A flow diagram of the Kawasaki Heavy Industries oxidation-absorption process is shown in Figure 2.3.10-1. Flue gas is taken from the boiler after the air preheater. It passes countercurrent to a magnesium hydroxide [$Mg(OH)_2$] slurry in the first section of the absorber. There, $SO_2$ is absorbed and undergoes the following reactions.[95]

$$SO_2(g) \rightarrow SO_2(aq) \tag{2-23}$$

$$Mg(OH)_2(s) + SO_2(aq) + 5H_2O \rightarrow MgSO_3 \cdot 6H_2O(s)\downarrow \tag{2-44}$$

The $SO_2$-free flue gas passes to the first denitrification section of the absorber while the liquid effluent drops to a holding tank. A recycle $NO_2$ stream is added to the flue gas to bring the $NO$:$NO_2$ mole ratio to 1:1. The resulting mixture then passes countercurrent to a $Mg(OH)_2$ slurry. Equimolar amounts of NO and $NO_2$ react and are reabsorbed in the following manner.[96]

$$NO(g) + NO_2(g) \rightarrow N_2O_3(g) \tag{2-45}$$

$$N_2O_3(g) \rightarrow N_2O_3(aq) \tag{2-46}$$

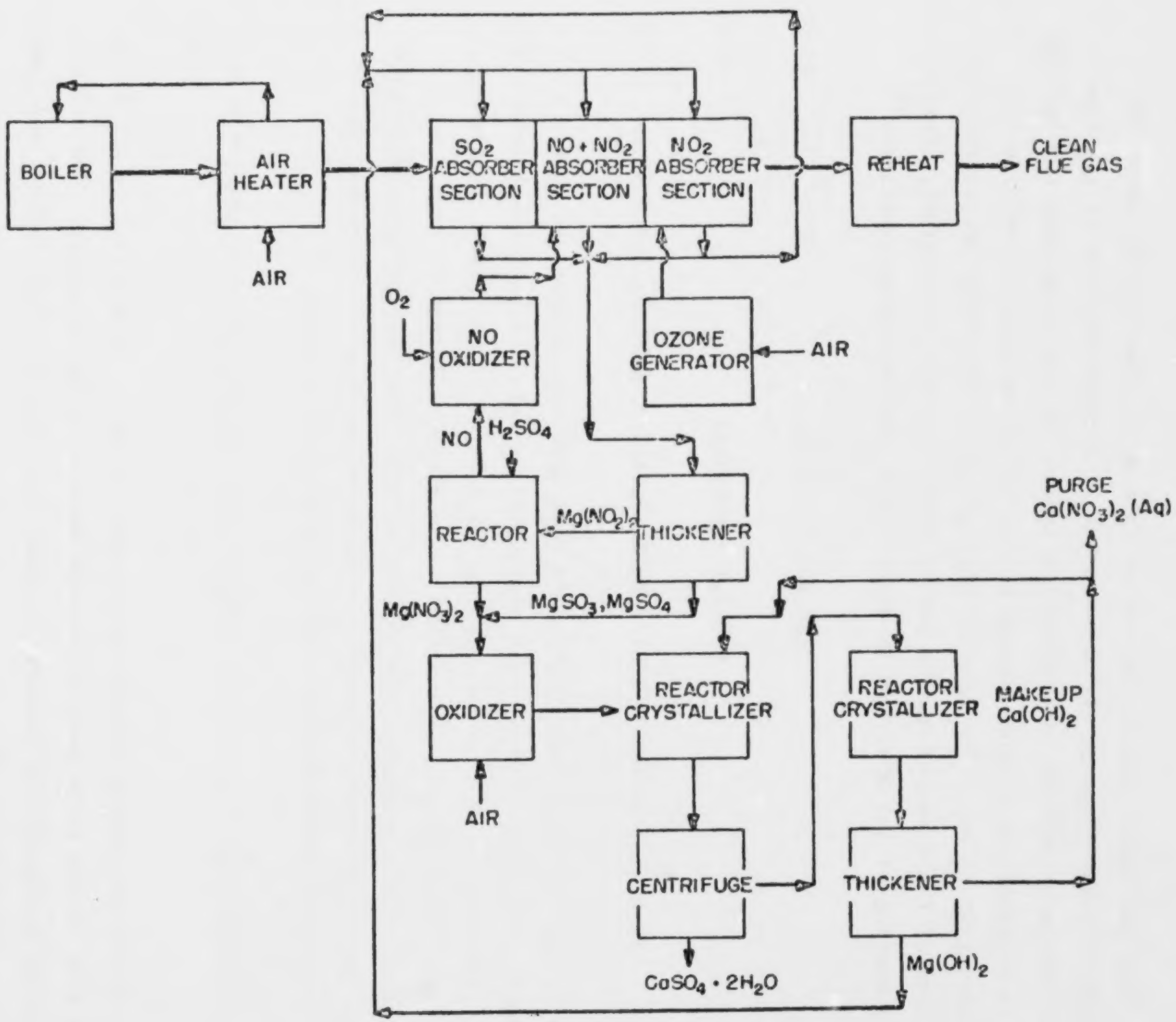

BOILER
AIR HEATER
AIR
$SO_2$ ABSORBER SECTION
NO + $NO_2$ ABSORBER SECTION
$NO_2$ ABSORBER SECTION
REHEAT
CLEAN FLUE GAS
$O_2$
NO OXIDIZER
OZONE GENERATOR
AIR
NO
$H_2SO_4$
REACTOR
$Mg(NO_2)_2$
THICKENER
PURGE
$Ca(NO_3)_2$ (Aq)
$Mg(NO_3)_2$
$MgSO_3, MgSO_4$
OXIDIZER
REACTOR CRYSTALLIZER
REACTOR CRYSTALLIZER
MAKEUP $Ca(OH)_2$
AIR
CENTRIFUGE
THICKENER
$CaSO_4 \cdot 2H_2O$
$Mg(OH)_2$

Figure 2.3.10-1. Flow diagram of Kawasaki Heavy Industries process.[94]

$$Mg(OH)_2(aq) + N_2O_3(aq) \rightarrow Mg(NO_2)_2(aq) + H_2O \qquad (2\text{-}47)$$

The flue gas passes out of the top of this absorption section while the liquid effluent drops to the holding tank. Because the rate of reaction 2-45 decreases with $NO_x$ concentration (below 200 ppm it becomes negligible), it is necessary to further reduce $NO_x$ by injecting ozone to oxidize the remaining NO to $NO_2$. The mixture then passes to the final denitrification section of the absorber and is passed countercurrent to a $Mg(OH)_2$ slurry. This section of the absorber is described by the following reactions.[97]

$$2NO_2(g) \rightarrow N_2O_4(g) \qquad (2\text{-}8)$$

$$N_2O_4(g) \rightarrow N_2O_4(aq) \qquad (2\text{-}9)$$

$$2N_2O_4(aq) + 2Mg(OH)_2(s) \rightarrow Mg(NO_3)_2(aq) + Mg(NO_2)_2(aq) + 2H_2O \qquad (2\text{-}48)$$

The clean flue gas leaves the top of this absorber section, is passed to a reheater for plume buoyancy and sent to the stack. Part of the liquid effluent from this section is recycled to the tops of the absorber sections while the rest drops to the holding tank. The slurry solution is pumped to a thickener which separates the soluble nitrite ($NO_2^-$) and nitrate ($NO_3^-$) salts from the solid magnesium sulfite. The overflow from the thickener passes to a $NO_2^-$ decomposition reactor where sulfuric acid is added.[98]

$$3Mg(NO_2)_2(aq) + 2H_2SO_4(aq) \rightarrow 2MgSO_4(aq) + Mg(NO_3)_2(aq) + 4NO(g)\uparrow + 2H_2O \qquad (2\text{-}49)$$

The NO off-gas passes through an oxidizer where it is oxidized by air to $NO_2$ and sent to the first denitrification section of the absorber. The effluent from the decomposition reactor is mixed with the thickener bottoms and pumped to a second oxidizer.[99]

$$MgSO_3\cdot6H_2O(s) + \tfrac{1}{2}O_2(g) \rightarrow MgSO_4(aq) + 6H_2O \qquad (2\text{-}50)$$

The magnesium sulfate formed in the oxidizer is treated with calcium nitrate in a gypsum production reactor.[25]

$$Ca(NO_3)_2(aq) + MgSO_4(aq) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \downarrow + Mg(NO_3)_2(aq) \quad (2\text{-}51)$$

The products of this reaction are sent to a centrifuge to remove the solid gypsum by-product. The liquid from the centrifuge goes to another decomposition reactor where makeup lime slurry is added.[100]

$$Mg(NO_3)_2(aq) + Ca(OH)_2(s) \rightarrow Ca(NO_3)_2(aq) + Mg(OH)_2(s) \quad (2\text{-}52)$$

The magnesium hydroxide product is separated in a thickener and recycled to the absorbers. The thickener overflow stream is split and part is recycled to the gypsum production reactor and the rest is concentrated to form low-grade liquid fertilizer by-product, $Ca(NO_3)_2$.

Since the processes in this category are all very different, especially with respect to chemistry, generalization of typical ranges of operating variables is not meaningful and, therefore, not presented. No published economics for these processes were found.

Presently, the equimolar absorption-type oxidation-absorption processes are still in the pilot-unit stage of development. Table 2.3.10-1 presents a list of all oxidation-absorption process vendors and their project's status of development. These processes have not yet been applied to oil-fired boilers.

TABLE 2.3.10-1. PROCESS VENDORS OF OXIDATION-ABSORPTION PROCESSES[100]

| Vendor | Status of Development |
|---|---|
| Kawasaki Heavy Industries | 1975: 5000 $Nm^3$/hr flue gas from coal-fired boiler |
| Tokyo Electric-MHI ($NO_x$ only) | 1974: 100,000 $Nm^3$/hr flue gas from natural gas-fired boiler |
| Ube Industries | No information available |

2.3.10.2 System Performance--

No oil-fired tests have been performed. Very little information has been published on any of the tests conducted.

The production of nitrate salts poses a potential secondary pollution problem. The plan for reclaiming and concentrating the nitrates as $Ca(NO_3)_2(aq)$ for liquid fertilizer is questionable as the by-product is of low quality and may not be easily marketable in the U.S. Also, the gypsum by-product would be contaminated with various nitrate and sulfite salts, and therefore, would probably be useful only as landfill material. Much corrosion-resistant material is necessary due to the utilization of ozone and circulating magnesium slurries. The three absorber sections, with their respective operating conditions, and ozone generation present complex process control problems. The process steps of several absorber sections in series (large fan requirements), ozone generation (corona discharge), flue gas reheat (inline heater), and by-product and wastewater treatment are all energy intensive and present technical and economic disadvantages when compared to other simpler FGT processes.

## 2.4 CONTROLS FOR NATURAL GAS-FIRED BOILERS

In the previous two sections which discuss controls for coal and oil-fired boilers, up to ten different process types are presented. Many of these process types are not considered here for application to natural gas-fired boilers for two reasons. First, natural gas-fired boilers have no $SO_2$ emission problems, and therefore, the simultaneous systems are not considered. Second, these boilers have no particulate emissions, and as a result, the systems designed specifically for high particulate applications are not considered. This leaves two systems to be considered for application to natural gas-fired boilers and these are discussed in the following sections.

### 2.4.1 Selective Catalytic Reduction-Fixed Packed Bed Reactor

#### 2.4.1.1 System Description--

Fixed packed bed systems are applicable only to flue gas streams containing less than 20 mg/$Nm^3$ of particulates. As such, they are applicable to natural gas-fired boilers.

The primary feature of these systems is the reactor which contains the catalyst. As the name implies, the granular catalyst is randomly packed in a stationary bed. An example of a typical fixed bed reactor is shown in Figure 2.4.1-1. The important features of the reactor are:

- the catalyst
- the catalyst support
- the gas distributor

The catalyst can be either spherical or cylindrical in shape. Spherical granules typically range in size from 4-10 mm in diameter.[103] The composition varies from process to process and most formulations are proprietary. The catalyst is supported either by inert packing (as shown in Figure 2.4.1-1) or by a perforated support plate (Figure 2.4.1-2). The catalyst supports

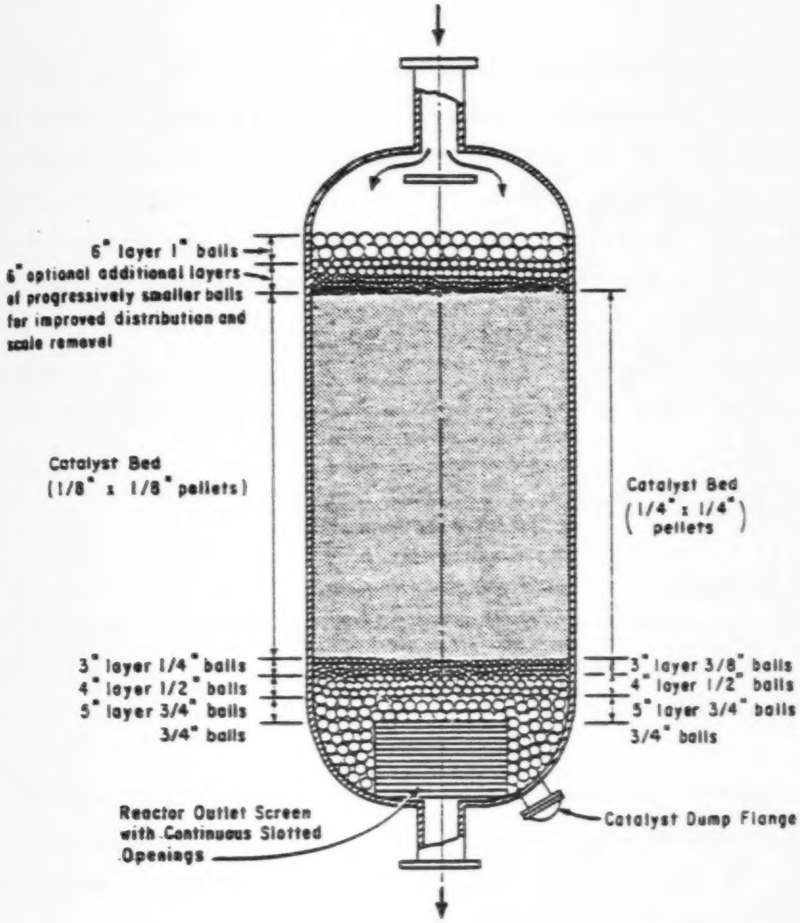

6" layer 1" balls
6" optional additional layers of progressively smaller balls for improved distribution and scale removal
Catalyst Bed (1/8" x 1/8" pellets)
Catalyst Bed (1/4" x 1/4" pellets)
3" layer 1/4" balls
4" layer 1/2" balls
5" layer 3/4" balls
3/4" balls
3" layer 3/8" balls
4" layer 1/2" balls
5" layer 3/4" balls
3/4" balls
Reactor Outlet Screen with Continuous Slotted Openings
Catalyst Dump Flange

Figure 2.4.1-1. Example of typical fixed packed bed reactor.[101]

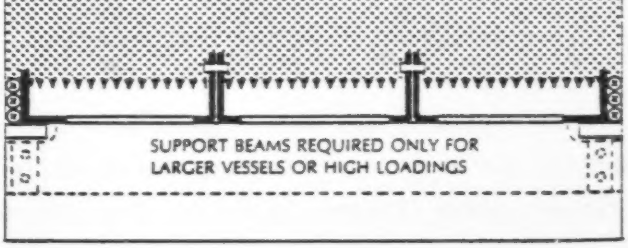

SUPPORT BEAMS REQUIRED ONLY FOR LARGER VESSELS OR HIGH LOADINGS

Figure 2.4.1-2. Example of catalyst support plate.[102]

hold the catalyst fixed in place in order to prevent both mobilization of the particles by the gas stream and catalyst rearrangement which would allow channelling of the flue gas. The gas distributor can be a perforated plate or similar device which spreads the gas flow across the entire cross-section of the catalyst bed.

A typical fixed bed SCR process layout is presented in Figure 2.4.1-3. Several arrangements are possible, however, for application to new boilers this arrangement is the most desirable.

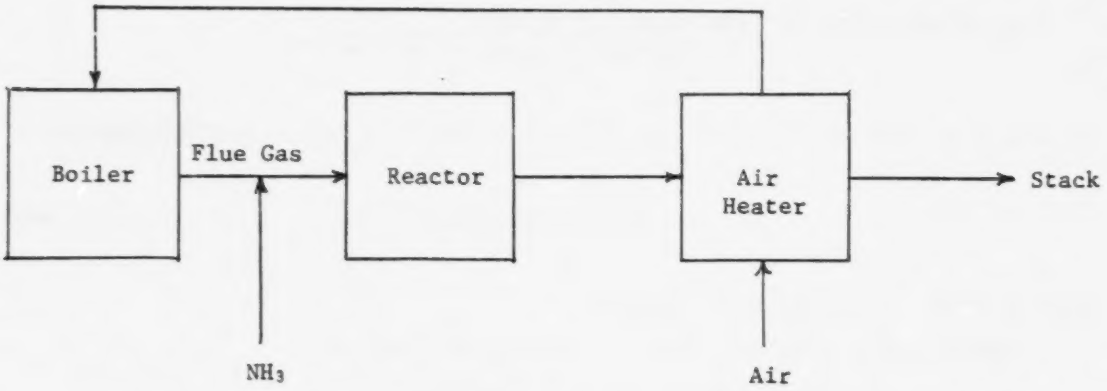


Figure 2.4.1-3. Process layout for fixed bed SCR process.

The principle of operation of these systems involves a gas phase reaction between ammonia ($NH_3$) and $NO_x$ (NO and $NO_2$). These reactions are presented most accurately by[12]

$$4NH_3 + 4NO + O_2 \rightleftarrows 4N_2 + 6H_2O \quad (2\text{-}53)$$

$$4NH_3 + 2NO_2 + O_2 \rightleftarrows 3N_2 + 6H_2O \quad (2\text{-}54)$$

The first reaction predominates since flue gas $NO_x$ is typically 90-95 percent NO. As shown, the $NO_x$ is reduced to molecular nitrogen ($N_2$) which exits with the flue gas stream.

The primary design equation used with these processes is the standard equation for reactor design,[13] represented by

$$\frac{V}{F} = \int_{o}^{x} \frac{dx}{r} \qquad (2\text{-}3)$$

where V is the catalyst volume

F is the mass (or molar) flow rate

x is the conversion of $NO_x$ to $N_2$

r is the reaction rate $\frac{\text{mass (or moles)}}{\text{volume of catalyst x time}}$

The reaction rate, r, for each NO reduction reaction can be represented by

$$r = k[NH_3]^a[NO]^b[O_2]^c \qquad (2\text{-}4)$$

where k is the reaction rate constant

$[NH_3]$, $[NO]$, $[O_2]$ are the reactant concentrations

a, b, c, are empirically determined exponents

The reaction rate is different for each catalyst formulation and, therefore, values for k, a, b, and c must be determined for the particular catalyst to be used before any design can be performed. The reaction rate constant is usually described by the Arrhenius equation

$$k = Ae^{-\frac{E}{RT}} \qquad (2\text{-}5)$$

where A is the frequency factor

E is the activation energy

R is the universal gas constant, and

T is the temperature

Values for k, a, b and c for two catalyst formulations are shown in Table 2.4.1-1. Values for other catalyst formulations will be different.

TABLE 2.4.1-1. REACTION RATE DATA FOR TWO CATALYST FORMULATIONS[11]

Catalyst: $V_2O_5$ on $Al_2O_3$

$$k = 2.05 \times 10^3 e^{-\frac{9650}{RT}}$$

$a = 0.30$

$b = 0.22$

$c = 0.05$

Catalyst: Fe-Cr on $Al_2O_3$

$$k = 3.25 \times 10^3 e^{-\frac{10,860}{RT}}$$

$a = 0.45$

$b = 0.10$

$c = 0.15$

The most important design and operating variables are similar to those for moving bed systems using granular catalysts. These are listed, along with typical ranges, in Table 2.4.1-2.

TABLE 2.4.1-2. DESIGN AND OPERATING VARIABLES FOR FIXED PACKED BED SYSTEMS[14]

| Variable | Typical Range<br>(For Oil) | <br>(For Gas) |
|---|---|---|
| Gas Velocity, m/s | 0.5 - 1.5 | 0.5 - 1.5 |
| Bed Depth, m | 0.2 - 0.6 | 0.2 - 0.4 |
| Space Velocity, $hr^{-1}$ | 6,000 - 10,000 | 8,000 - 15,000 |
| Pressure Drop, $mmH_2O$ | 40 - 80 | 40 - 70 |
| Temperature, °C | 350 - 400 | 300 - 400 |

Other variables that affect the process are:

- flue gas flow rate
- $NO_x$ control level
- $NO_x$ concentration
- boiler load variation

The flue gas flow rate and $NO_x$ control level determine the catalyst volume required (hence reactor size). Increases in either parameter also increase the reactor size. The $NO_x$ concentration is primarily a function of fuel type used in the standard boilers. Higher concentrations require larger $NH_3$ storage and vaporization equipment; reactor size is not affected. Boiler load can affect several things including flue gas temperature, flow rate and $NO_x$ concentration. It is usually necessary to maintain reaction temperatures of 350 to 400°C. Temperature control equipment may be necessary to accomodate large boiler load variations which cause lower flue gas temperatures. Where these variations are present, some equipment overdesign may be warranted to insure a constant control level. These variables are discussed in more detail in the section on moving bed SCR systems for coal-fired boilers, Section 2.2.2. Costs of fixed packed bed systems range from $16-49/kW (capital) and 1.2-1.8 mills/kWh (operating). These costs are based on utility applications as well as a variety of process and operating conditions.

There are vendors of fixed packed bed SCR systems and all are Japanese. Vendors are listed in Table 2.4.1-3 and the scale of development is also noted. Fixed packed systems have been applied to industrial and utility boilers in Japan. Existing installations are shown in Tables 2.4.1-4 and 2.4.1-5. Currently, there are no installations in the U.S.

### 2.4.1.2 System Performance--

Typical performance data for fixed packed bed SCR systems are shown in Figures 2.4.1-4 through 2.4.1-8. These data indicate that $NO_x$ removals 90 percent and higher are achievable with these systems. This allows them to be considered for all control levels of interest in this study.

TABLE 2.4.1-3. VENDORS OF SCR FIXED BED SYSTEMS FOR GAS-FIRED APPLICATIONS[21]

| Vendor | Notes |
|---|---|
| Sumitomo | Tested on commercial scale equipment |
| Hitachi Zosen | Tested on commercial scale equipment |
| Hitachi, Ltd. | Tested on commercial scale equipment |
| Mitsubishi Heavy Industries | Tested on commercial scale equipment |
| Ishikawajima-Harima Heavy Industries | Tested on commercial scale equipment |
| Mitsui Toatsu Chemical | Has not been to boilers |
| Kawasaki Heavy Industries | Tested on pilot scale equipment |
| Mitsubishi Kakoki Kaisha | Tested on commercial scale equipment |

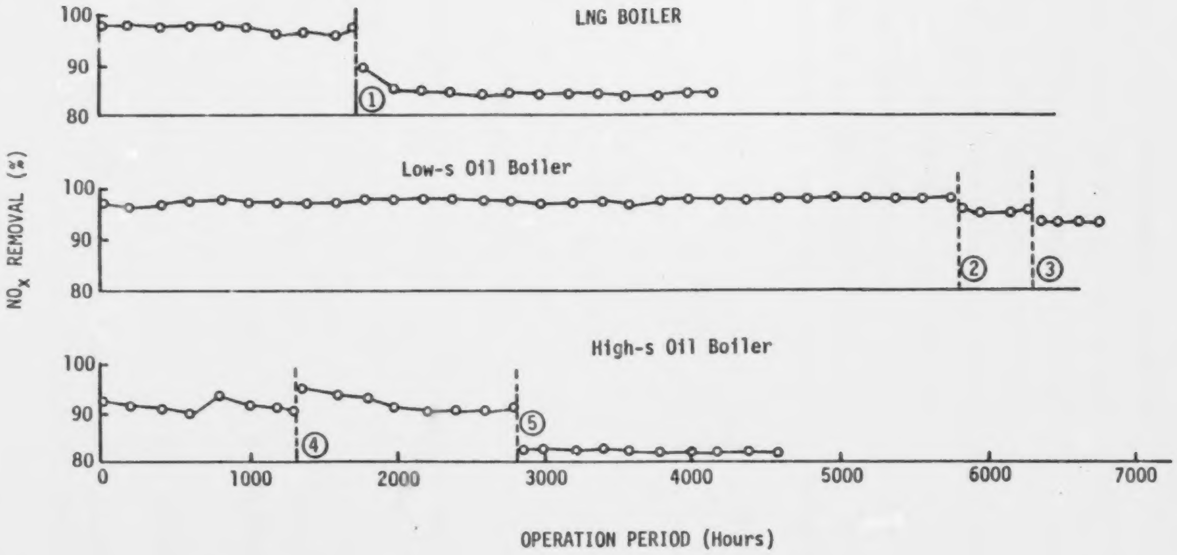
TABLE 2.4.1-4. EXISTING FCT INSTALLATIONS OF SCR FIXED BED SYSTEMS GAS-FIRED INDUSTRIAL BOILERS[21]

| Location | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Takaishi | Osaka Gas | Mitsubishi H.I. | LNG | 30,000 | December 1976 |

TABLE 2.4.1-5. EXISTING FGT INSTALLATIONS OF SCR FIXED BED SYSTEMS GAS-FIRED UTILITY BOILERS[21]

| Location | User | Process Developer | Fuel | Capacity ($Nm^3/hr$) | Completion Date |
|---|---|---|---|---|---|
| Kokura | Kyushu Electric | Mitsubishi H.I. | LNG | 3,380,000* | October 1978 |
| Chita | Chubu Electric | Hitachi, Ltd. | LNG | 4,000,000* | April 1978 |

*Flow rate is combined value from two boilers.



Circled figures show times when SV and $NH_3/NO$ mole ratio were changed.

1. SV 10,000 20,000 $hr^{-1}$
2. SV 10,000 15,000 $hr^{-1}$
3. SV 15,000 20,000 $hr^{-1}$
4. SV 6,200 4,500 $hr^{-1}$
5. SV 4,500 6,200 and the mole ratio 0.95 0.83

Figure 2.4.1-4. Test results at gas- and oil-fired boilers.[124]

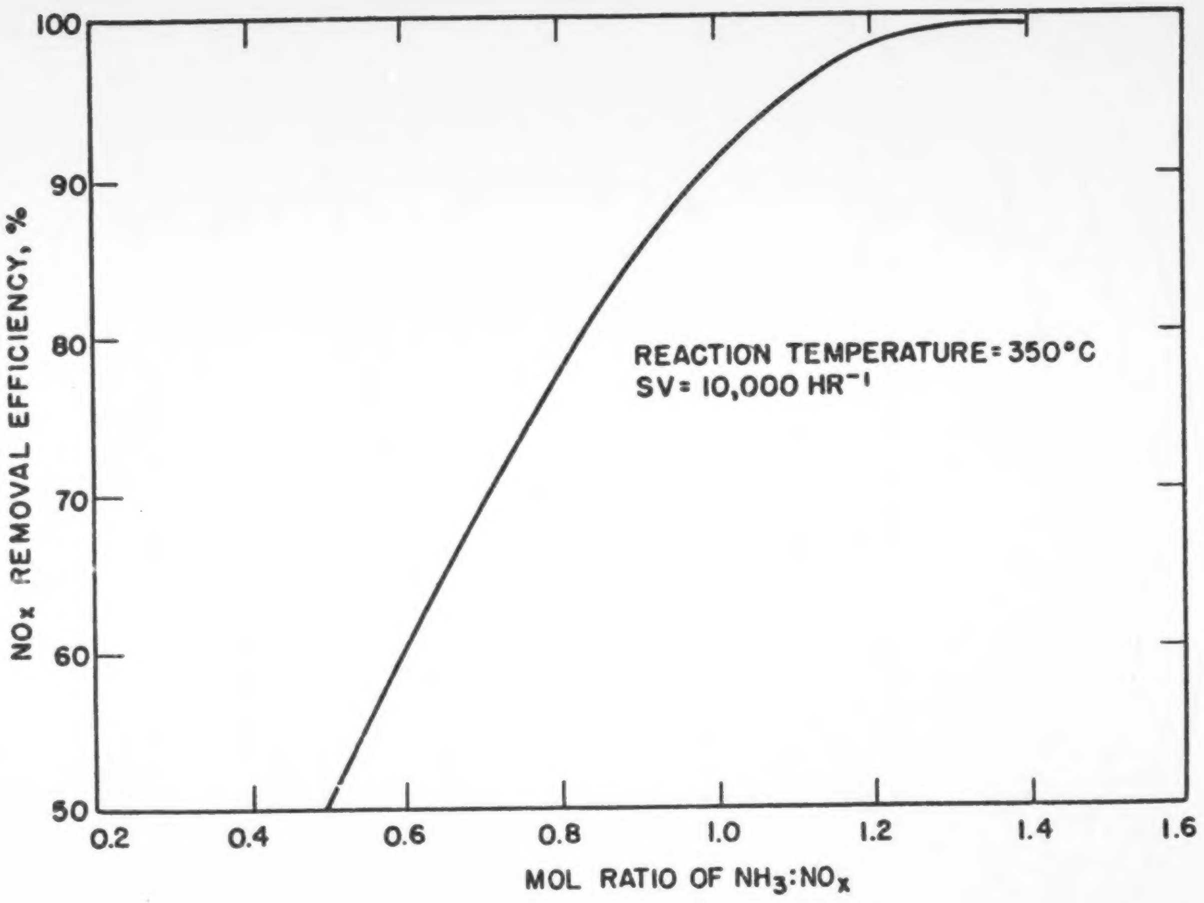


Figure 2.4.1-5. Characteristic curve of the effect of mol ratio of $NH_3$:$NO_x$ on $NO_x$ removal efficiency for Hitachi, Ltd. Process.[125]

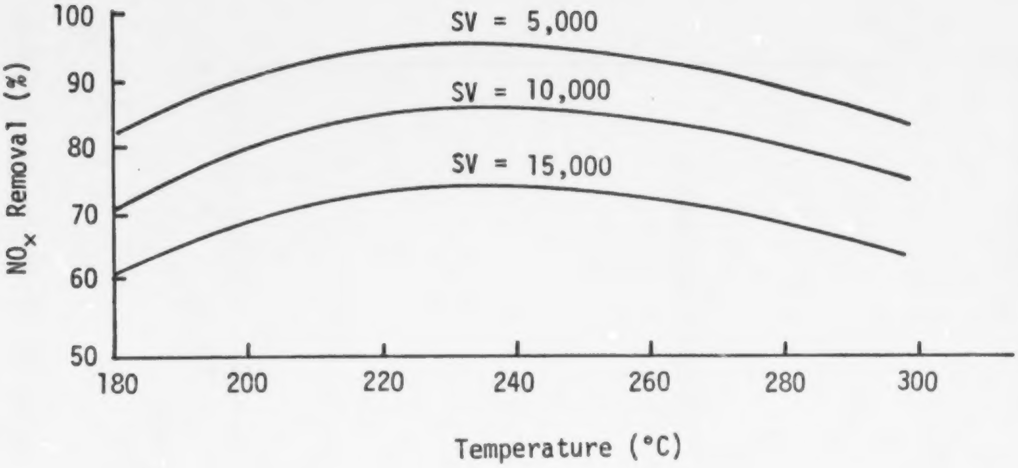


Figure 2.4.1-6. Performance of catalyst MTC-102 (flue gas by LPG burning).[126]

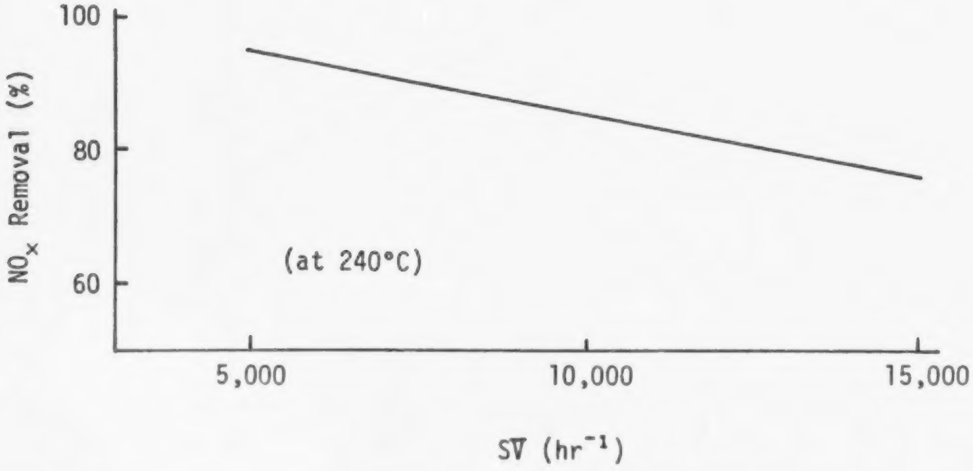


Figure 2.4.1-7. SV and $NO_x$ removal (MTC-102) (flue gas by LPG burning).[127]

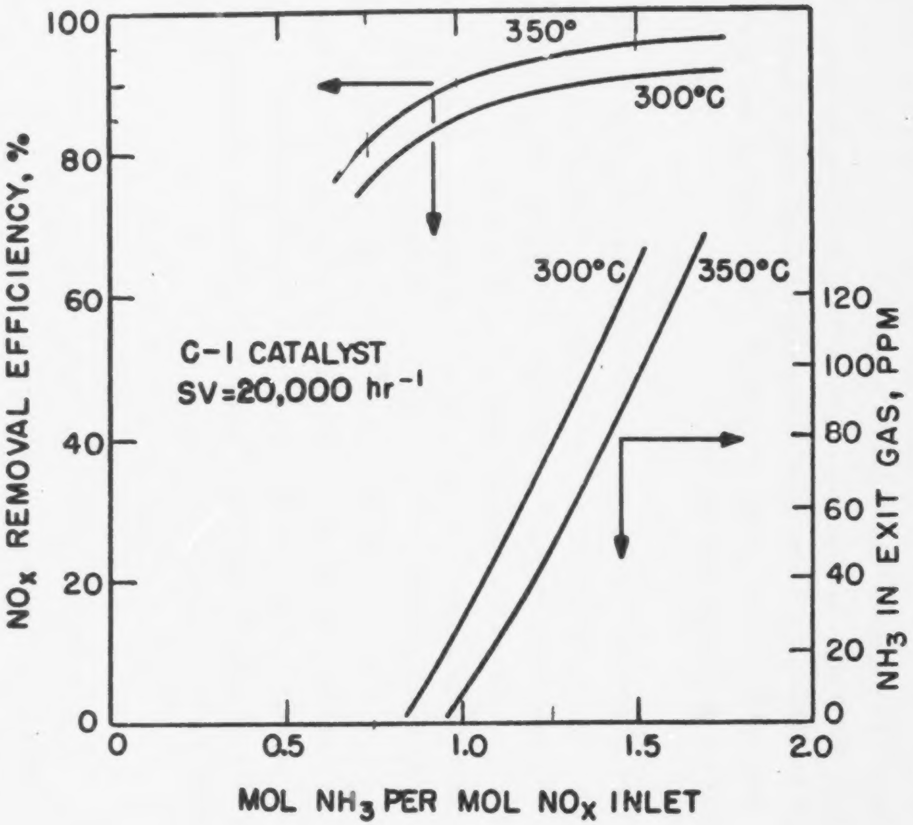


Figure 2.4.1-8. Relationship among inlet $NH_3$:$NO_x$ mol ratio, $NO_x$ removal efficiency, and exiting $NH_3$ concentration using the Sumitomo Chemical C-1 Catalyst.[128]

### 2.4.2 Absorption-Oxidation

#### 2.4.2.1 System Description--

Absorption-oxidation processes remove $NO_x$ from flue gas by absorbing the NO or $NO_x$ into a solution containing an oxidant which converts the $NO_x$ to a nitrate salt. Two types of gas/liquid contactors can be used and examples of each type are shown in Figure 2.4.2-1. Both perforated plate and packed towers accomplish $NO_x$ absorption by generating high gas/liquid interfacial areas. The choice of one type of contactor is a design decision made to achieve a given removal for the least cost.

A generalized process flow diagram is shown in Figure 2.4.2-2. Flue gas is taken from the boiler after the air preheater. Before the gas can be sent to the $NO_x$ absorber, it must be $SO_2$-free since $SO_2$ consumes prohibitive amounts of the costly liquid-phase oxidant. This is not a problem with natural gas fired-boilers since they have no $SO_2$ emissions. In most cases, the oxidant is permanganate ($MnO_4^-$). The flue gas enters the distributing space at the bottom of the $NO_x$ absorber, below the packing or plates. The gas passes upward through the column, countercurrent to the flow of the liquid absorbent/oxidant (usually a KOH solution containing $KMnO_4$). $NO_x$ is absorbed and then oxidized over the length of the column according to the following reactions.[31]

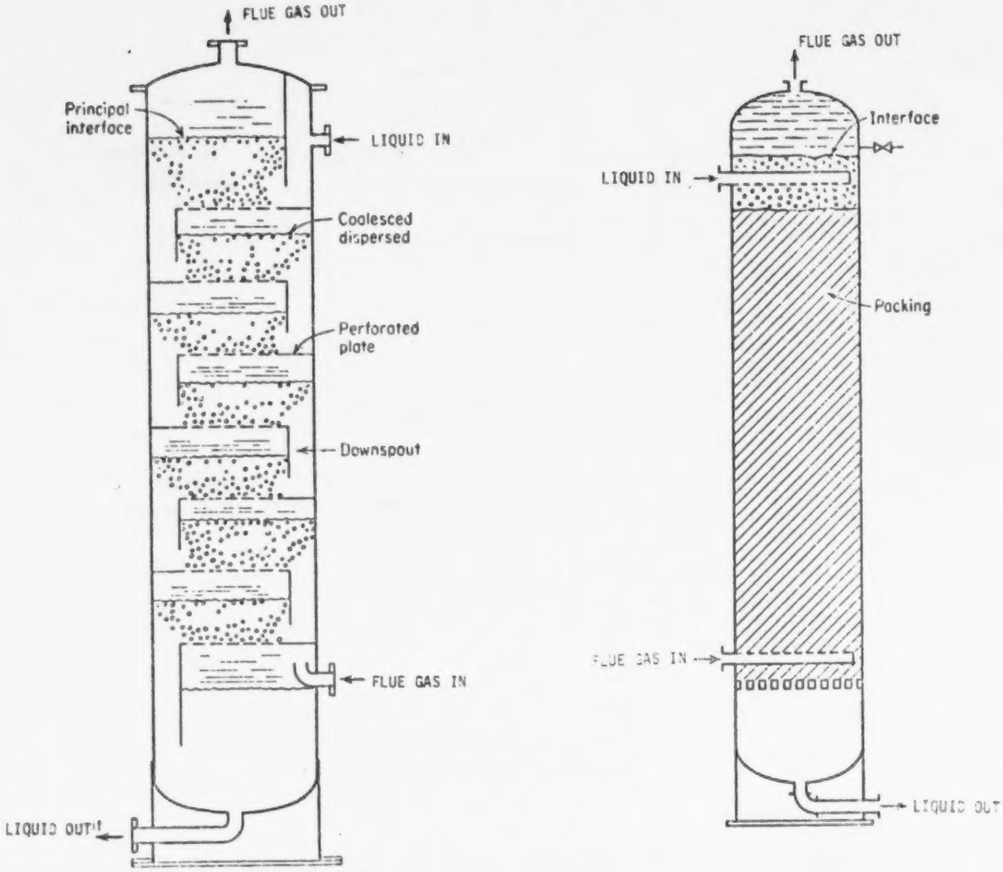
$$NO(g) \rightarrow NO(aq) \qquad (2\text{-}6)$$

$$NO(aq) + KMnO_4(aq) \rightarrow KNO_3(aq) + MnO_2(s)\downarrow \qquad (2\text{-}7)$$

$$2NO_2(g) \rightarrow N_2O_4(g) \qquad (2\text{-}8)$$

$$N_2O_4(g) \rightarrow N_2O_4(aq) \qquad (2\text{-}9)$$

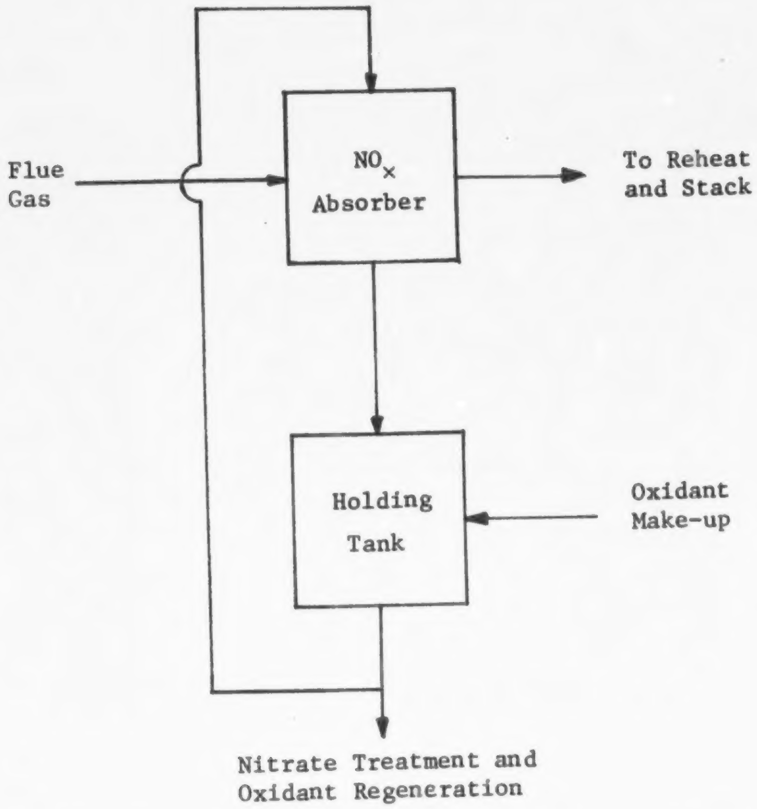
$$N_2O_4(aq) + 2K_2MnO_4(aq) \rightarrow 2KMnO_4(aq) + 2KNO_2(aq) \qquad (2\text{-}10)$$


FLUE GAS OUT
Principal interface
LIQUID IN
Coalesced dispersed
Perforated plate
Downspout
FLUE GAS IN
LIQUID OUT
FLUE GAS OUT
Interface
LIQUID IN
Packing
FLUE GAS IN
LIQUID OUT

Perforated Plate Absorber

Packed Absorber

Figure 2.4.2-1. Gas/liquid contactor options for Absorption-Oxidation Processes.[29]



Flue Gas
$NO_x$ Absorber
To Reheat and Stack
Holding Tank
Oxidant Make-up
Nitrate Treatment and Oxidant Regeneration

Figure 2.4.2-2. Process flow diagram for absorption-oxidation process.[30]

Since most of the $NO_x$ from combustion processes occurs as NO,[32] reactions 2-6 and 2-7 predominate. The clean gas passes out of the top of the absorber to a heater for plume buoyancy and is sent to the stack. The absorbing solution drops to a holding tank where makeup KOH and/or $KMnO_4$ are added. This solution flows to a centrifuge to separate the solid $MnO_2$ which is then electrolytically oxidized to $MnO_4$. The remaining solution is either concentrated in an evaporator to form a weak $KNO_3$ solution or is electrochemically treated to produce a weak $HNO_3$ solution and a mixed stream of KOH and $KNO_3$.

The fundamental design equation used for gas absorption column design is

$$\int_{Y_a}^{Y_b} \frac{dy}{(y-y^*)} = \left(\frac{K_y a}{G_y}\right) Z \qquad (2\text{-}11)$$

where $y$ = bulk $NO_x$ concentration (mole fraction) of gas phase at any given point in column

$y-y^*$ = overall driving force for absorption ($y^*$ being the $NO_x$ concentration of a gas in equilibrium with a given liquid $NO_x$ concentration)

$Y_b$ = inlet $NO_x$ concentration

$Y_a$ = outlet $NO_x$ concentration

$K_y$ = overall gas-phase mass transfer coefficient, lb-moles $NO_x$/(ft$^2$)(hr)(mole fraction)

$a$ = area of gas-liquid interface per unit packed volume, ft$^2$/ft$^3$

$G_y$ = molal gas mass velocity, lb-moles flue gas/(ft$^2$)(hr)

$Z$ = length of packed section of column, ft

In a column containing a given packing or plate configuration and being irrigated with a certain liquid flow, there is an upper limit to the gas flow rate. This limit's superficial gas velocity (volumetric gas flow rate/cross-sectional area of column) is called the flooding velocity. At this point, the gas flow completely impedes the downward motion of the liquid and blows the liquid out of the top of the column. The gas velocity, obviously, must be lower than the flooding velocity. How much lower is a design decision. Often, it is an economic tradeoff between power costs and equipment costs. A low gas velocity will lower the pressure drop and, hence, the power costs but the absorber will have a larger diameter and cost more. High gas velocities have an opposite effect. Usually the optimum gas velocity is about one-half the flooding velocity.[34] The height of the column depends on the desired level of removal and on the rate of mass transfer. The latter is a major problem for these systems trying to achieve large $NO_x$ reductions since NO is relatively insoluble in water. This can be seen in Table 2.4.2-1.

TABLE 2.4.2-1. NITROGEN OXIDES CHARACTERISTICS[35]

| | Boiling Point, °C | Solubility in Cold Water (0°C), $cm^3$ | Solubility in Hot Water (60°C), $cm^3$ |
|---|---|---|---|
| NO | -151.8 | 7.34/100 cc $H_2O$ | 2.37/100 cc $H_2O$ |
| $NO_2$ | 21.2 | soluble, decomposes | -- |

One can see that NO has a very limited solubility in water and, since most $NO_x$ is present as NO, the rate of mass transfer (absorption) is going to be relatively slow. This means that the absorber must be tall with a high liquid flow rate. Table 2.4.2-2 presents the effects of boiler/flue gas variables on the design of absorption-oxidation systems.

TABLE 2.4.2-2. SYSTEM DESIGN CONSIDERATIONS

| Variable | Design Effect |
|---|---|
| Presence of particulates | Requires prescrubber |
| Presence of $SO_2$ | Requires FGD pretreatment |
| Increased gas flow | Requires larger column diameter; increased liquid flow rate |
| Increased $NO_x$ concentration | Requires larger column height; increased oxidant concentration |

Both flue gas flow rate and $NO_x$ concentration can be affected by boiler operating conditions. Therefore a change in load on an industrial boiler may alter these variables markedly. The absorber must be designed to accommodate any anticipated load changes. The column size and the liquid and oxidant flows must be designed for each application after examining the boiler operating history and establishing ranges of variation.

None of the sources consulted for this study could supply typical ranges for operating variables such as liquid/gas ratio, reagent concentrations or pressure drops and, as a result, none are presented here. Economic data were not presented either. One source did estimate the removal for absorption-oxidation processes to be 85 percent.[36]

Presently, absorption-oxidation processes are still in the pilot unit stage of development. Table 2.4.2-3 presents a list of absorption-oxidation process vendors and the status of development of their projects. One can see from the table that no gas-fired flue gas tests have been performed.

TABLE 2.4.2-3. PROCESS VENDORS OF ABSORPTION-OXIDATION PROCESSES[37,38]

| Vendor | Status of Development |
|---|---|
| Hodogaya | No information available; stopped development on process |
| Kobe Steel | 1974: 1000 $Nm^3$/hr gas from iron-ore sintering furnace; stopped development on process |
| MON (Mitsubishi Metal, MKK, Nikon Chemical) | 1974: 4000 $Nm^3$/hr flue gas from oil-fired boiler |
| Nissan Engineering | 1972: 4 pilot plants, 100-2000 $Nm^3$/hr tail gas from $HNO_3$ plant |

### 2.4.2.2 System Performance--

No gas-fired tests have been made. No information has been published on tests conducted with other fuels. The relative insolubility of NO in water may present a major obstacle to achieving the stringent level of control (90 percent $NO_x$ reduction) by absorption-oxidation processes. Another primary drawback of absorption-oxidation systems is the production of nitrate salts (see Equation 2-7), a secondary pollutant. These processes probably could not be applied on a large scale as wastewater treatment systems (chemical or biological) do not remove nitrogen compounds from the wastewater.[39] Trying to recover the nitrates as nitric acid for industrial use or potassium nitrate for fertilizer does not seem promising as the by-products are of low quality. Also, the use of an expensive, liquid-phase oxidant requires stainless steel and other corrosion resistant materials of construction. The process steps of oxidant regeneration (electrolysis) and flue gas reheat (inline heater) are all energy intensive and present technical and economic disadvantages.

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80. *Ibid.*, pp. 32, 40, 87, 104, 139.

81. *Ibid.*, p. 94.

82. *Ibid.*, p. 389.

83. *Ibid.*, p. 93.

84. *Ibid.*, p. 95.

85. Ando, Jumpei. October 1978, *op cit.*, p. 7-12.

86. Faucett. H. L., *op cit.*, p. 96.

87. Ando, Jumpei. October 1978, *op cit.*, p. 7-14.

88. Faucett, H. L., *op cit.*, pp. 63, 99.

89. *Ibid.*, pp. 42-44, 63-66, 98-100, 117-121, 125-130.

90. Audo, Jumpei. October 1978, *op cit.*, p. 7-25.

91. Faucett, H. L., *op cit.*, p. 25.

92. *Ibid.*, pp. 52, 61, 97, 116, 126.

93. Ando, Jumpei. October 1978, *op cit.*, p. 110.

94. Faucett, H. L., *op cit.*, p. 70.

95. *Ibid.*, p. 69.

96. *Ibid.*, p. 71.

97. *Ibid.*, p. 71.

98. *Ibid.*, p. 72.

99. *Ibid.*, p. 72.

100. *Ibid.*, pp. 73, 146, 149.

101. Rase, Howard F. Chemical Reactor Design for Process Plants. Volume 1. Wiley-Interscience. 1977. p. 515.

102. *Ibid.*, p. 514.

103. Ando, Jumpei. October 1978, *op cit.*, p. 3-30.

104. *Ibid.*, p. 4-10.

105. *Ibid.*, p. 3-7.

106. *Ibid.*, p. 4-21.

107. *Ibid.*, p. 4-71.

108. *Ibid.*, p. 4-5.

109. *Ibid.*, p. 4-37.

110. *Ibid.*, p. 4-94.

111. *Ibid.*, p. 4-93.

112. *Ibid.*, p. 4-92.

113. *Ibid.*, p. 4-126.

114. *Ibid.*, p. 4-21.

115. *Ibid.*, p. 4-96.

116. *Ibid.*, p. 3-45.

117. *Ibid.*, p. 4-41.

118. Faucett, H. L., *op cit.*, p. 224

119. Ando, Jumpei. October 1978, *op cit.*, p. 4-95.

120. *Ibid.*, p. 6-29.

121. *Ibid.*, p. 7-31.

122. *Ibid.*, p. 7-19.

123. *Ibid.*, p. 7-20.

124. *Ibid.*, p. 4-43.

125. Faucett, H. L., *op cit.*, p. 214.

126. Ando, Jumpei. October 1978, *op cit.*, p. 4-121.

127. *Ibid.*, p. 121.

128. Faucett, H. L., *op cit.*, p. 298.

129. Noblett, J.G., *et al.* "Impact of $NO_x$ Selective Catalytic Reduction Processes on Flue Gas Desulfurization Processes," Draft Final Report EPA Contract No. 68-02-2608, Task 70, Radian Corporation, September 1979.

# SECTION 3

# CANDIDATES FOR BEST SYSTEMS OF EMISSION REDUCTION

The ten systems discussed in Section 2 are not applicable to all combinations of boiler types and fuels of interest in this study. However, several of these systems may be applicable to a specific boiler/fuel combination (*i.e.*, capable of removing sufficient $NO_x$ to meet proposed emission regulations). In this section, $NO_x$ control techniques which are applicable to the various boilers and fuels considered in this study are selected. The section is organized to compare $NO_x$-only and simultaneous $NO_x/SO_x$ reduction systems separately. The result is a set of candidate control techniques that will be evaluated in detail in subsequent sections to determine the "best" system for $NO_x$ control by FGT.

## 3.1 CRITERIA FOR SELECTION

Two sets of evaluation criteria are used to determine the set of candidate systems. One is the level of $NO_x$ control desired which determines the set of systems available for further evaluation. The other is a set of evaluation criteria that will allow comparison of the systems capable of meeting a particular level of control.

### 3.1.1 Factors Considered in Selection of Best Systems

A consistent set of rating criteria was used to evaluate and compare each of the FGT systems described in section 2 that are capable of achieving the proposed $NO_x$ removal levels. These criteria and the weighting factors are shown in Table 3.1.1-1. As can be seen, the criteria receiving most emphasis are status of development, economics, performance, and reliability.

TABLE 3.1.1-1. RATING CRITERIA AND WEIGHTING FACTORS

| Evaluation Category | Total Points |
|---|---|
| Performance | 14 |
| Operational/Maintenance Impacts on Performance | 7 |
| Preliminary Environmental Impacts | 9 |
| Preliminary Economic Impacts | 15 |
| Preliminary Energy/Material Impacts | 10 |
| Boiler Operation and Safety | 4 |
| Reliability | 14 |
| Status of Development | 16 |
| Adaptability to Existing Sources | 6 |
| Compatability with Other Control Systems | 5 |
| | 100 |

Emphasis is placed on the most developed FGT systems since they represent the most likely controls to be applied if a high degree of $NO_x$ control is required on industrial boilers. An FGT system must achieve the necessary $NO_x$ reduction and do so as economically as possible, hence the heavy emphasis on performance and economics. These are important considerations for any application. Reliability is heavily weighted because it is common for an industrial boiler to supply one or several continuous manufacturing processes. A high reliability is required to avoid frequent boiler shutdowns with subsequent loss of revenues due to dependency of the manufacturing process on the boiler.

It should be pointed out that only large differences in point values are significant while small differences are not. For example, ratings which differ by a factor of two are significant. However, two ratings 10 points apart do not necessarily indicate the superiority of one process. A more detailed breakdown of the evaluation criteria and the point values assigned is presented in Table 3.1.1-2. The basis for the detailed breakdown is discussed below. The analysis of each system using these criteria is discussed in Section 3.2.

TABLE 3.1.1-2. SPECIFIC POINT VALUES ASSOCIATED WITH SELECTION FACTORS

| Item | Quality | Points |
|---|---|---|
| 1. Performance | | |
| a. Desired control level (stringent, intermediate, or moderate) as percent of system's maximum design capability $\left(\frac{\text{Desired Control Level}}{\text{Maximum Design Control Level}}\right) \times 100$ | <70 | 8 |
| | 70 - 80 | 6 |
| | 80 - 90 | 4 |
| | 90 -100 | 2 |
| | >100 | No Go |
| b. Particulate handling capability | Great | 4 |
| | Some | 1 |
| | None | 0 |
| c. Load following ability | Good | 2 |
| | Fair | 1 |
| | Poor | 0 |
| 2. Operation and Maintenance Impacts on Performance | | |
| a. Moving parts | Few | 1 |
| | Many | 0 |
| b. Solids handling | No | 1 |
| | Yes | 0 |
| c. Process separability | Once-through | 1 |
| | Regenerable | 0 |
| d. Flue gas composition sensitivity | No | 1 |
| | Yes | 0 |
| e. Prescrubbing necessary | No | 1 |
| | Yes | 0 |
| f. Process stability | Simple process & insensitive control needs | 2 |
| | Complex process or sensitive control needs | 1 |
| | Complex process & sensitive control needs | 0 |
| 3. Preliminary Environmental Impacts | | |
| a. Secondary pollutants - Air | None | 3 |
| | Potential | 2 |
| | Some | 1 |
| | Major | 0 |
| - Liquid | None | 3 |
| | Some | 1 |
| | Major | 0 |

TABLE 3.1.1-2. (Continued)

| Item | Quality | Points |
|---|---|---|
| 3. a. Secondary pollutants (Cont'd) | | |
| - Solid | None | 3 |
| | Some | 1 |
| | Major | 0 |
| 4. Preliminary Economic Impacts | | |
| a. Capital investment | <50% mean | 7 |
| | 50% mean | 6 |
| | 75% mean | 5 |
| | Mean | 4 |
| b. Operating costs | 125% mean | 3 |
| | 150% mean | 2 |
| | >150% mean | 1 |
| c. Marketable by-product | Potential | 1 |
| | None | 0 |
| 5. Preliminary Energy/Material Impacts | | |
| a. Electrical demand | <1% output | 5 |
| | 1 - 2% | 4 |
| | 2 - 3% | 3 |
| | 3 - 4% | 2 |
| | 4 - 5% | 1 |
| | >5% | 0 |
| b. Auxiliary fuel use | No | 1 |
| | Yes | 0 |
| c. Energy intensive regeneration or by-product treatment | None | 2 |
| | Some | 1 |
| | Heavy | 0 |
| d. Raw material demand | Light | 2 |
| | Moderate | 1 |
| | Heavy | 0 |
| 6. Boiler Operation and/or Safety | | |
| Boiler impacts or safety hazards | None | 4 |
| | Potential | 2 |
| | Yes | 0 |
| 7. Reliability | | |
| a. Plugging and scaling | None | 5 |
| | Some | 2 |
| | Much | 0 |

TABLE 3.1.1-2. (Continued)

| Item | Quality | Points |
|---|---|---|
| 7. b. Simplicity - Number process steps | <3 | 6 |
| | 3 | 5 |
| | 4 | 4 |
| | 5 | 3 |
| | 6 | 2 |
| | 7 | 1 |
| | >7 | 0 |
| c. Material of construction | Carbon steel | 3 |
| | Some corrosion resistant material | 1 |
| | Much corrosion resistant material | 0 |
| 8. Development Status | | |
| a. Scale demonstrated | Commercial | 10 |
| | Prototype | 8 |
| | Pilot | 5 |
| | Bench | 2 |
| | Conceptual | 0 |
| b. Length of operation | >5000 hours | 3 |
| | 3000 - 5000 | 2 |
| | 1000 - 3000 | 1 |
| | <1000 | 0 |
| c. Uncertainties in technology | No | 3 |
| | Yes | 0 |
| 9. Adaptability to Existing Sources | | |
| a. Retrofit | Easy | 3 |
| | Difficult | 0 |
| b. Land required | Small | 3 |
| | Large | 0 |
| 10. Compatability with Other Control Systems | | |
| a. FGD | Yes | 3 |
| | No | 0 |
| b. ESP, other | Yes | 2 |
| | No | 0 |

#### 3.1.1.1 Performance--

A primary concern in the selection of an $NO_x$ flue gas treatment system is the system's performance. The first aspect to consider here is the $NO_x$ removal capability. This study is organized by different levels of $NO_x$ control (stringent, intermediate, moderate). The processes' maximum removal capability is compared to these various control levels to show the ease with which the system can meet the removal requirement. Another measurement of a system's performance is its load following capability--how well the system responds to a sudden change in boiler load. Generally, large, complex systems do not respond to load changes as quickly as small, simple systems. Slow response is a disadvantage since it may result in increased emissions during load changes.

#### 3.1.1.2 Operational and Maintenance Impacts--

This category is important for several reasons. A system with difficult operational steps or high maintenance requirements is not as desirable since it will require more manpower and increase operating costs. Reliability may also be adversely affected. For most FGT systems, this type of data is not available. In this study these impacts are inferred by examining each system and applying engineering judgment. The more mechanically complex a system is, the more likely it is to have operation and maintenance problems.

#### 3.1.1.3 Preliminary Environmental Impacts--

This category, along with the economic and energy categories, relies on published information for data. Detailed analyses of the candidate systems in these areas will be conducted in a subsequent section. The data presented in this section are used for comparison purposes only. Obviously it is undesirable for an FGT system to remove $NO_x$ at the expense of emitting a secondary pollutant. For this reason secondary pollutants (air, liquid, and solid) emitted by the process, or potentially so, are identified. Systems with no secondary pollutants receive the highest ratings.

### 3.1.1.4 Preliminary Economic Impacts--

With an industrial boiler it is probable that application of FGT will affect the price of products from a new or modified facility and thereby affect the salability of these products. For this reason, the lowest cost system that will adequately control $NO_x$ is desirable. The areas considered are capital investment ($/kW), operating costs (mills/kWh), and credits for marketable by-products. Cost data that are available are primarily for utility installations. While there is some economy of scale in the investment cost due to the large size of the facilities, the values are adequate for preliminary cost comparisons. Sample economy of scale calculations showing how the preliminary economic figures were generated are contained in Appendix II.

### 3.1.1.5 Preliminary Energy/Material Impacts--

It is desired to minimize energy and raw material consumption by an FGT process since this also minimizes operating costs. In addition, dependence on outside factors such as raw material supplies is reduced. The main system parameters considered are the electrical demand of the system, use of auxiliary fuels and energy, and intensive regeneration or by-product treatment processes. Also, heavy raw material demands are noted. Again, utility data are used for comparative purposes since very little industrial boiler data are available.

### 3.1.1.6 Boiler Operation and/or Safety--

It is desirable to minimize impacts of the FGT system on the boiler. The main areas of potential impacts are air heater fouling, duct scaling and stack corrosion. These impacts as well as safety aspects of the process are determined by inspection of the process equipment and chemistry.

### 3.1.1.7 Reliability--

Reliability data are not generally available for all of the process types considered. Many have not been applied on commercial scale equipment. Some reliability data are available for SCR systems, but data from other systems are necessary before the reliability of SCR systems can be compared on a relative basis. For most systems it can be said that simplicity is concommitant with reliability and this concept is used in the evaluation.

### 3.1.1.8 Development Status--

A crucial consideration in the selection of the best $NO_x$ control techniques by flue gas treating is the status of development of the processes. Presently, there are but a few commercial-size $NO_x$ FGT units in operation on industrial boilers--all in Japan. Because most of the flue gas treatment development work has been conducted fairly recently, it is vital that those systems which have been demonstrated most fully be given primary consideration for implementation to industrial boilers. For this study, availability by the year 1981 was estimated using the current status of development and reported on-going development. The size of the unit, length of operation, and any uncertainties in technology were all taken into account.

### 3.1.1.9 Adaptability to Existing Sources--

Since applying FGT to modified existing sources is generally more difficult than with new sources, the ease of retrofit was examined. Structural and equipment modifications necessary for retrofit are considered since existing boilers are not constructed to accommodate FGT systems. Land requirements of the FGT system are also considered, since existing industrial boilers are not necessarily located near large land areas. Quite frequently, they are located in the center of a plant and surrounded by equipment. Small systems requiring little boiler modification are desired.

### 3.1.1.10 Compatibility with Other Control Systems--

This category is related to retrofit and new installation. Where additional control equipment is existing or planned for installation, an FGT system which does not affect and is not affected by other control systems is desirable. This aspect of the processes is determined by inspection of the chemistry and equipment of the FGT system as well as other pollutant control systems.

### 3.1.2 Selection of Control Levels--Moderate, Stringent, and Intermediate

The control levels selected are applied to the following boilers:

| Fuel | Type | Load ($MW_t$) | Uncontrolled NO$_x$ Emissions | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | (lb/hr) | | | (lb/$10^6$ Btu) | | |
| Gas | Firetube | 4.4 | 2.63 | | | 0.18 | | |
| Gas | Watertube | 44 | 26.26 | | | 0.18 | | |
| Oil-dist. | Firetube | 4.4 | 2.38 | | | 0.16 | | |
| Oil-dist. | Watertube | 44 | 23.76 | | | 0.16 | | |
| Oil-resid. | Watertube | 44 | 60.00 | | | 0.40 | | |
| Oil-resid. | Watertube | 8.8 | 12.00 | | | 0.40 | | |
| | | | High S Eastern | Low S Eastern | Low S Western | High S E | Low S E | Low S W |
| Coal | Underfeed Stoker | 8.8 | 19.05 | 16.35 | 23.40 | 0.64 | 0.55 | 0.78 |
| Coal | Chaingrate | 22 | 47.70 | 40.80 | 58.65 | 0.64 | 0.54 | 0.78 |
| Coal | Spreader Stoker | 44 | 95.40 | 81.45 | 117.15 | 0.64 | 0.54 | 0.78 |
| Coal | Pulverized Coal | 58.6 | 152.46 | 130.50 | 187.56 | 0.76 | 0.65 | 0.94 |

These NO$_x$ emission levels are all lower than the following average State Implementation Plan (SIP) requirements except for one oil-fired boiler, one coal-fired boiler burning high sulfur eastern coal, and all coal-fired boilers burning low sulfur western coal.

| Fuel | $NO_x\left(\frac{lb}{10^6\ Btu}\right)$ |
|---|---|
| Coal | 0.7 |
| Oil | 0.3 |
| Gas | 0.2 |

The moderate level of control is defined as representing that level which is achievable applying techniques in current practice within industry. This is the least stringent emission reduction achievable applying accepted engineering practice. For FGT systems, this represents an $NO_x$ removal of approximately 70 percent. When considering $NO_x$ FGT, it is not reasonable to consider a removal level less than 70% since such levels can probably be achieved by combustion modification techniques at lower costs. Allowable $NO_x$ emissions at this control level are shown below:

| Fuel | Emission Level $\left(\frac{lb\ NO_x}{10^6\ Btu}\right)$ |
|---|---|
| Coal | 0.24 |
| Oil | 0.09 |
| Gas | 0.06 |

Most of the control techniques are capable of controlling the standard boilers with the highest $NO_x$ emissions at this level.

The stringent level of control is defined as a technology-forcing level and represents the most rigorous control which might be considered. This represents an $NO_x$ removal of 90 percent. Allowable emissions at this control level are shown below:

| Fuel | Emission Level $\left(\frac{lb\ NO_x}{10^6\ Btu}\right)$ |
|---|---|
| Coal | 0.08 |
| Oil | 0.03 |
| Gas | 0.02 |

These systems are operating at their upper limit of practical $NO_x$ removal capability to achieve this level of control and are definitely technology-forcing.

Intermediate level of control is defined as a level between moderate and stringent and probably representing a technological or cost breakpoint. At this point in time, it is difficult to say if those logical breakpoints exist and, if so, where they are. Therefore, the intermediate level was chosen between moderate and stringent levels. The intermediate levels of control considered here represent about 80 percent $NO_x$ removal. Allowable emissions at this level are shown below:

| Fuel | Emission Level $\frac{\text{lb } NO_x}{10^6 \text{ Btu}}$ |
|---|---|
| Coal | 0.16 |
| Oil | 0.06 |
| Gas | 0.04 |

The best FGT systems should be able to achieve steady-state control at this level. This control level provides an alternative choice between the least-stringent and technology-forcing options.

The allowable emission rates for each of the control levels are summarized in Table 3.1.2-1.

TABLE 3.1.2-1. CONTROLLED EMISSION LEVELS IN THIS STUDY (lb/$10^6$ Btu)

| | Moderate | Intermediate | Stringent |
|---|---|---|---|
| Coal | 0.24 | 0.16 | 0.08 |
| Oil | 0.09 | 0.06 | 0.03 |
| Gas | 0.06 | 0.04 | 0.02 |

## 3.2 BEST CONTROL SYSTEMS FOR COAL-FIRED BOILERS

A three phase selection process was used to determine the best $NO_x$ control systems. The first phase involves comparing the maximum removal level obtainable by each process with the level of control desired--moderate, intermediate or stringent. Those process types which cannot achieve this level are eliminated from further consideration. The remaining process types are then evaluated using the criteria established in Section 3.1.1. The result is a set of process types that are most desirable for a particular consideration of special characteristics of the process types in the set in order to determine the best system candidates. For example, all SCR processes may rate high for application to gas-fired boilers. However, the SCR fixed packed bed process may be more applicable than the moving bed or parallel flow SCR processes since ability to tolerate particulates is not required for gas-fired boiler applications.

For use in the application of the selection factors, tables are compiled which list the process features pertinent to each selection factor. The data in these tables was derived from information presented in Section 2. For coal-fired boilers, this information is presented in Tables 3.2-1 and 3.2-2.

### 3.2.1 Moderate Reduction Controls

The first phase evaluation eliminated the adsorption process from consideration since it cannot achieve 70% $NO_x$ reduction at high $NO_x$ concentrations (400-600 ppm). Application of the selection factors resulted in numerical ratings for the remaining processes as shown in Table 3.2.1-1. As can be seen, the four SCR processes were superior. The fixed packed bed technique was eliminated since it would rapidly plug due to the high particulate levels encountered with coal-fired applications. Therefore, the candidate systems for moderate control of coal-fired boilers are SCR parallel flow and SCR moving bed for $NO_x$-only removal and SCR parallel flow for simultaneous $NO_x/SO_x$ removal.

## TABLE 3.2-1. COMPARISON INFORMATION OF $NO_x$-ONLY SYSTEMS FOR COAL-FIRED BOILERS

| | Performance | Operational and maintenance impacts | Preliminary environmental impacts | Preliminary economic impacts | Preliminary energy and material impacts |
|---|---|---|---|---|---|
| SCR Fixed Packed Bed | Capable of attaining >90% $NO_x$ control level; cannot be used with high particulate levels; good load following capability. | Few moving parts; gas phase chemistry; simple process - good controllability; need high removal of particulates - ESP; large pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates:<br>Capital: \$130/kW[1]<br>Operating:<br>2.1 mills/kWh[1,2]<br>Cost is higher than other SCR's due to ESP. | Electrical usage: 1.2% of total output; large $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio); may require auxiliary heater. |
| SCR Moving Bed | Capable of attaining >90% $NO_x$ control level; can be used with some particulates (up to 2.0 $g/Nm^3$); adequate load following capability. | Moving parts, solids handling - increased maintenance; gas phase chemistry; fairly simple - controllable; need particulate removal; low pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates:<br>Capital: \$92/kW[3]<br>Operating: 2.0 mills/kWh | Electrical usage: unknown - should be <1%; large $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio); may require auxiliary heater; greater catalyst attrition due to moving bed. |
| SCR Parallel Flow | Capable of attaining >90% $NO_x$ control level; can be used with full particulate loading (up to 20 $g/Nm^3$); good load following capability. | Few moving parts; gas phase chemistry; simple process - good controllabiltiy; moderate pressure drop; no particulate removal needed. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates:<br>Capital: \$44/kW[4]<br>Operating: 1.5 mills/kWh | Electrical usage: 0.2% of total output; large $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio); may require auxiliary heater. |
| Absorption-Oxidation | No removal data are available - should be able to achieve moderate control level; can be used with full particulate loading; fair load following capability. | Complex process with sensitive control needs; sensitive to flue gas sulfur content - separate $SO_x$ scrubber before $NO_x$ absorber; prescrubber needed to remove particulates and $Cl^-$; very large pressure drops. | $NO_3^-$ salts in wastewater. | 20 MW estimates: none available, but since process contains extra scrubber train,<br>Capital: \$500/kW<br>Operating: 8 mills/kWh | Electrical usage: unknown, estimate ≈3%; uses large amounts of gas-phase oxidant and by-product treatment materials. |

TABLE 3.2-1. (Continued)

| | Boiler operation and/or safety | Reliability | Status of development | Adaptability to existing sources | Compatibility with other control systems |
|---|---|---|---|---|---|
| SCR Fixed Packed Bed | No safety hazards. | Catalyst easily plugged; possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | Has only been tested on bench-scale (8 oil- and numerous gas-fired commercial operations). | Some difficulty; few pieces of process equipment; little land needed. | Excessive reheat required if after FGD; needs ESP. |
| SCR Moving Bed | No safety hazards. | Little catalyst plugging; possible $NH_4HSO_4$ scaling; fairly simple - few process steps; little corrosion resistant material. | Has only been tested on bench-scale (5 oil- and 3 coke oven gas-fired commercial operations). | Some difficulty; few pieces of process equipment; little land needed. | Excessive reheat required if after FGD; needs particulate removal. |
| SCR Parallel Flow | No safety hazards. | Little catalyst plugging (must be packed well); possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | Has only been tested on bench-scale; pilot plants due to start up in 1979 (some oil-fired operations); commercial operation by 1981. | Some difficulty; catalyst can be placed in duct between economizer and preheater without a separate reactor; few pieces of equipment; little land needed. | Completely compatible with FGD. |
| Absorption-Oxidation | Oxidant handling could be hazardous. | Numerous process steps and corrosion resistant material. | Has not been tested on coal-fired flue gas (a few pilot plants treating oil-fired and furnace gases). | Much land needed for numerous pieces of process equipment and wastewater treatment. | Existing FGD would be helpful as process cannot tolerate sulfur. |

## TABLE 3.2-2. COMPARISON INFORMATION OF SIMULTANEOUS $NO_x/SO_x$ SYSTEMS FOR COAL-FIRED BOILERS

| | Performance | Operational and maintenance impacts | Preliminary environmental impacts | Preliminary economic impacts | Preliminary energy and material impacts |
|---|---|---|---|---|---|
| SRC Parallel Flow | Capable of attaining 90% control of both $NO_x$ and $SO_x$; can be applied to gases with high particulate loadings; process can follow boiler load easily through use of gas bypass arrangement. | Process has several sections but all except $NO_x/SO_x$ reactor are based on well established technology; average maintenance requirements. | Potential $NH_3$ emissions. | 20 MW estimates:<br>Capital: $475/kW<br>Operating: 5 mills/kWh | Electrical usage: 1.5% of total output; also consumes $NH_3$, naphtha, and steam. |
| Adsorption | Capable of attaining 60% $NO_x$ control level; cannot be used with high particulate levels; poor load following capability; primarily $SO_x$ removal. | Many moving parts, hot solids handling; complex process; need ESP for particulate removal; major maintenance requirements; high pressure drop. | Ash disposal. | 20 MW estimates:<br>Capital: $215 kW[5]<br>Operating: 2.3 mills/kWh | Electrical usage: unknown - should be ≈2%; activated char usage high due to attrition. |
| Electron Beam Radiation | Capable of attaining 80% $NO_x$ control level; cannot be used with particulates; fair load following capability; also removes $SO_x$. | Simple process but complex control; sensitive to flue gas composition (at least 1% $O_2$ and $H_2O > NO_x$). | $H_2SO_4$ mist and a powder containing ammonium nitrates and sulfates are generated. | 20 MW estimates:<br>Capital: $202 kW[6]<br>Operating: unknown<br>Electricity is only major. | Electrical usage: 3.3% of total output (excluding ESP); treatment of by-product is unknown. |
| Absorption-Reduction | Capable of attaining 85% $NO_x$ control level; can be used with full particulate loading; good load following capability; removes $SO_x$. | Complex process with very sensitive control needs; sensitive to flue gas composition (low $O_2$ and $SO_x$:$NO_x$ ratio >2.5); need prescrubber to remove particulates and $Cl^-$. | Possibility of plume from absorbent (sulfate or $NH_3$). | 20 MW estimates:<br>Capital: $413/kW[7]<br>Operating: 7.4 mills/kWh<br>Gypsum by-product (landfill). | Electrical usage: 1.8% of total output; large amounts of chelating compound, absorbent, and regeneration chemicals are used. |
| Oxidation-Absorption-Reduction | Capable of attaining 90% $NO_x$ control level; can be used with full particulate loading; poor load following capability - oxidant generation lagtime; removes $SO_x$. | Complex process with very sensitive control needs; prescrubber needed to remove particulates and $Cl^-$; large pressure drop. | $NO_3^-$ or N-S salts or $NH_3^-$ based compounds in wastewater. | Economic estimates: unknown for coal-fired plant; gypsum by-product (landfill). | Electrical usage: 9.0% of total output; uses large amounts of gas-phase oxidant and by-product treatment materials. |
| Oxidation-Absorption | Capable of attaining 90% $NO_x$ control level; can be used with full particulate loading; poor load following capability; removes $SO_x$. | Complex process with very sensitive control needs; prescrubber needed; large pressure drop. | $NO_3^-$ salts in wastewater. | Economic estimates: unknown; gypsum by-product and liquid fertilizer, or $HNO_3$. | Electrical usage: unknown (will be ≈10% of total output): uses large amounts of gas-phase oxidant and by-product treatment materials. |

TABLE 3.2-2. (Continued)

| | Boiler operation and/or safety | Reliability | Status of development | Adaptability to existing sources | Compatibility with other control systems |
|---|---|---|---|---|---|
| SCR Parallel Flow | $H_2$ usage may present safety hazard. | Process steps well established; should be reliable. | $SO_2$ system has been tested on coal-fired flue gas; $NO_x/SO_x$ operation with coal-fired flue gas to begin late 1979; pilot unit tests; $SO_2$ work up and $H_2$ generation not tested, but are established technology. | Will need land for equipment. | Compatible with particulate systems. |
| Adsorption | Possible safety hazard due to poor char distribution in beds. | Char plugged by particulates; numerous process steps; some corrosion resistant material in high temperature zones. | One prototype unit treating coal-fired flue gas. | Need land for pieces of process equipment. | Suitable for placement after ESP; not useful with FGD system as $NO_x$ removal is secondary. |
| Electron Beam Radiation | Radiation safety hazards are unknown as are those of byproduct. | Few process steps; stainless steel reactor. | Has not been tested on coal-fired flue gas (one pilot plant treating gas from sintering machine); uncertain by-product treatment method. | Need land for pieces of process equipment. | Needs ESP; with or without existing FGD but capital cost will be the same. |
| Absorption-Reduction | No safety hazards. | Many process steps; much glass- and elastomer-lined equipment. | Has not been tested on coal-fired flue gas (several pilot plants treating oil-fired flue gas); $NO_x$ absorption chemistry uncertain. | Much land needed for numerous pieces of process equipment. | Cannot be used in conjunction with FGD. |
| Oxidation-Absorption-Reduction | Gas-phase oxidant presents serious safety hazard. | Numerous process steps and corrosion resistant material; oxidant generation system subject to periodic failure. | Has not been tested on coal-fired flue gas (6 prototype units treating oil-fired flue gas in operation). | Much land needed for numerous pieces of process equipment, oxidant generation, and wastewater treatment. | Cannot be used in conjunction with FGD. |
| Oxidation-Absorption | Gas-phase oxidant presents serious safety hazard. | Numerous process steps and corrosion resistant material; oxidant generation system subject to periodic failure. | One pilot plant treating flue gas from coal-fired boiler. | Much land needed for numerous pieces of process equipment, oxidant generation, and wastewater treatment. | Compatible. |

TABLE 3.2.1-1. CANDIDATE SYSTEMS SELECTION: COAL-FIRED BOILERS - MODERATE CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| NO$_x$-Only | | | |
| SCR Fixed Packed Bed | 69 | no | Adversely affected by particulates |
| SCR Moving Bed | 70 | no | Adversely affected by particulates |
| SCR Parallel Flow | 83 | yes | |
| Absorption-Oxidation | 43 | no | Low rating |
| Simultaneous NO$_x$/SO$_x$ | | | |
| SCR Parallel Flow | 72 | yes | |
| Adsorption | NA | no | |
| Electron Beam Radiation | 41 | no | Low rating |
| Absorption-Reduction | 52 | no | Low rating |
| Oxidation-Absorption-Reduction | 51 | no | Low rating |
| Oxidation-Absorption | 51 | no | Low rating |

NA - Not applicable (see Appendix).

A detailed listing of how each process was evaluated on each selection factor is contained in Tables A3.1 and A3.2 in the Appendix.

### 3.2.2 Stringent Reduction Controls

In a similar manner, candidate systems for stringent control were selected. The results appear in Table 3.2.2-1. A detailed listing of the selection factors and point values for each system is contained in Tables A3.3 and A3.4. The candidate systems selected are SCR parallel flow and SCR moving bed for $NO_x$-only removal and SCR parallel flow for simultaneous $NO_x/SO_x$ removal.

### 3.2.3 Intermediate Reduction Controls

The selection results for this level are presented in Table 3.2.3-1. Detailed application of the selection factors is presented in Tables A3.5 and A3.6. The candidate systems selected are SCR parallel flow and SCR moving bed for $NO_x$-only removal and SCR parallel flow for simultaneous $NO_x/SO_x$ removal.

## 3.3 BEST CONTROL SYSTEMS FOR OIL-FIRED BOILERS

The control systems for oil-fired boilers were evaluated using the same method described in the previous section on coal-fired boilers. Tables 3.3-1 and 3.3-2 present a side-by-side comparison of all potential systems with data categorized with respect to the selection factors. The information in this table is summarized from Section 2. The table is similar in many respects to the equivalent table for coal. This is due to the fact that, since FGT systems are applied after the boiler, they are relatively insensitive to the types of fuel burned. Two notable exceptions are particulate and sulfur emissions which are a function of the fuel type. Process characteristics that change with fuel type are noted in the table.

TABLE 3.2.2-1. CANDIDATE SYSTEMS SELECTION: COAL-FIRED BOILERS - STRINGENT CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 62 | no | Adversely affected by particulates |
| SCR Moving Bed | 60 | no | Adversely affected by particulates |
| SCR Parallel Flow | 73 | yes | |
| Absorption-Oxidation | NA | no | |
| Simultaneous $NO_x$/$SO_x$ | | | |
| SCR Parallel Flow | 68 | yes | |
| Adsorption | NA | no | |
| Electron Beam Radiation | NA | no | |
| Absorption-Reduction | NA | no | |
| Oxidation-Absorption-Reduction | 48 | no | Low rating |
| Oxidation-Absorption | 49 | no | Low rating |

NA - Not applicable (see Appendix).

TABLE 3.2.3-1. CANDIDATE SYSTEMS SELECTION: COAL-FIRED BOILERS - INTERMEDIATE CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 67 | no | Adversely affected by particulates |
| SCR Moving Bed | 69 | no | Adversely affected by particulates |
| SCR Parallel Flow | 81 | yes | |
| Absorption-Oxidation | 43 | no | Low rating |
| Simultaneous $NO_x/SO_x$ | | | |
| SCR Parallel Flow | 70 | yes | |
| Adsorption | NA | no | |
| Electron Beam Radiation | 41 | no | Low rating |
| Absorption-Reduction | 50 | no | Low rating |
| Oxidation-Absorption-Reduction | 49 | no | Low rating |
| Oxidation-Absorption | 46 | no | Low rating |

NA - Not applicable (see Appendix).

## TABLE 3.3-1. COMPARISON INFORMATION OF $NO_x$-ONLY SYSTEMS FOR OIL-FIRED BOILERS

| | Performance | Operational and maintenance impacts | Preliminary environmental impacts | Preliminary economic impacts | Preliminary energy and material impacts |
|---|---|---|---|---|---|
| SCR Fixed Packed Bed | Capable of achieving >90% $NO_x$ reduction; cannot be used with high particulate levels; good load following capability. | Few moving parts; gas phase chemistry; simple process - good controllability; need high removal of particulates - ESP; large pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates:<br>Capital: $70/kW[8,9]<br>Operating:<br>1.9 mills/kWh[10,11] | Electrical usage: unknown for oil-fired plant; moderate $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio); may require auxiliary heater. |
| SCR Moving Bed | Capable of achieving >90% $NO_x$ reduction; can tolerate particulate level of most oils (<1 g/Nm³); adequate load following capability. | Some moving parts, solids handling - increased maintenance; gas phase chemistry; simple - controllable; low pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates:<br>Capital: $70/kW[11,12]<br>Operating: 1.8 mills/kWh | Electrical usage: unknown for oil-fired plant; moderate $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio); may require auxiliary heater; greater catalyst attrition due to moving bed. |
| SCR Parallel Flow | Capable of achieving >90% $NO_x$ reduction; can tolerate full particulate loading (up to 20 g/Nm³); good load following capability. | Few moving parts; gas phase chemistry; simple process - good controllability; no particulate removal needed; moderate pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates:<br>Capital: $39/kW[13]<br>Operating: unknown. | Electrical usage: unknown for oil-fired plant; moderate $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio); may require auxiliary heater. |
| Absorption-Oxidation | No removal data are available; can tolerate particulates; fair load following capability; removes $SO_x$. | Complex process with sensitive control needs; sensitive to flue gas sulfur content - separate $SO_x$ scrubber before $NO_x$ absorber; prescrubber needed; very large $\Delta P$. | $NO_3^-$ salts in wastewater. | Economic estimates: unknown. | Electrical usage: unknown; uses large amounts of liquid phase oxidant and regeneration materials. |

TABLE 3.3-1. (Continued)

| | Boiler operation and/or safety | Reliability | Status of development | Adaptability to existing sources | Compatibility with other control systems |
|---|---|---|---|---|---|
| SCR Fixed Packed Bed | No safety hazards. | Catalyst easily plugged; possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | 8 commercial operations in Japan. | Some difficulty; few pieces of process equipment; little land required. | Excessive reheat required if after FGD; needs ESP. |
| SCR Moving Bed | No safety hazards. | Some catalyst plugging; possible $NH_4HSO_4$ scaling; fairly simple - few process steps; little corrosion resistant material. | 6 commercial oil-fired operations in Japan. | Some difficulty; few pieces of process equipment; little land required. | Excessive reheat required if after FGD. |
| SCR Parallel Flow | No safety hazards. | Little catalyst plugging; possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | Numerous commercial oil-fired operations in Japan. | Some difficulty; catalyst can be placed in duct between economizer and preheater without a separate reactor; few pieces of equipment; little land needed. | Completely compatible with FGD. |
| Absorption-Oxidation | Oxidant handling can be dangerous. | Numerous process steps and corrosion resistant material. | 2 pilot plants treating oil-fired flue gas. | Much land needed for numerous pieces of process equipment and wastewater treatment. | Existing FGD would be helpful as process cannot tolerate sulfur. |

## TABLE 3.3-2. COMPARISON INFORMATION OF SIMULTANEOUS $NO_x/SO_x$ SYSTEMS FOR OIL-FIRED BOILERS

| | Performance | Operational and maintenance impacts | Preliminary environmental impacts | Preliminary economic impacts | Preliminary energy and material impacts |
|---|---|---|---|---|---|
| SCR Parallel Flow | Capable of attaining 90% control of both $NO_x$ and $SO_2$; can be used with full particulate loading; good load following capability. | Most of the process steps are based on well established technology; average maintenance requirements. | Potential $NH_3$ emissions. | Economic estimates: unknown for oil-fired plant; assumed to be similar to those for coal.<br>20 MW estimates for coal:<br>Capital: $475/kW<br>Operating: 5 mills/kWh | Electrical usage: unknown for oil-fired plant; assumed to be similar to those for coal, i.e. 1.5% of boiler output as electricity; also uses steam, naphtha and $NH_3$. |
| Adsorption | Capable of attaining 60% $NO_x$ reduction; cannot be used with high particulate levels; poor load following capability; primarily $SO_x$ removal. | Many moving parts, hot solids handling - high maintenance; complex process; may need particulate removal on residual oils; large pressure drop. | Ash disposal. | Ecomonic estimates: unknown for oil-fired plant; elemental S by-product. | Electrical usage: unknown for oil-fired plant; large activated char demand due to attrition. |
| Electron Beam Radiation | Capable of attaining 80% $NO_x$ removal; cannot be used with particulates; fair load following capability; also removes $SO_x$. | Simple process with complex control; sensitive to flue gas composition (at least 1% $O_2$ and $H_2O > NO_x$); may need particulate removal on residual oils. | $H_2SO_4$ mist and a powder containing ammonium-nitrates and sulfates are generated. | Economic estimates: unknown for oil-fired plant; electricity is primary operating expense. | Electrical usage: unknown for oil-fired plant; treatment of by-product is unknown. |
| Absorption-Reduction | Capable of attaining 85% removal; can tolerate particulates; good load following capability; cannot be used on distillate oil; also removes $SO_x$. | Complex process with very sensitive control needs; sensitive to flue gas composition (low $O_2$ and $SO_x$:$NO_x$ ratio >2.5); need prescrubber; large $\Delta P$. | Possibility of plume (sulfite or $NH_3$) from absorbent. | 20 MW estimates:<br>Capital: $187/kW[14,15]<br>Operating: 5.4 mills/kWh<br>Gypsum by-product (landfill). | Electrical usage: 1.8% of total output; extremely large amounts of chelating compound; absorbent and regeneration chemicals are used. |
| Oxidation-Absorption-Reduction | Capable of attaining 90% $NO_x$ reduction; can tolerate particulates; poor load following capability - oxidant generation lagtime; cannot be used on distillate oils; removes $SO_x$. | Complex process with very sensitive control needs; prescrubber needed; large pressure drop. | $NO_3^-$ or N-S salts or NH-base compounds in wastewater. | 20 MW estimates:<br>Capital: $231/kW[16,17]<br>Operating: 6.4 mills/kWh<br>Gypsum by-product (landfill). | Electrical usage: 5-10% of total output; large amount of gas phase oxidant and by-product treatment materials. |
| Oxidation-Absorption | Capable of attaining 90% $NO_x$ reduction; can tolerate particulates; poor load following capability; removes $SO_x$. | Complex process with very sensitive control needs; prescrubber needed; large pressure drop. | $NO_3^-$ salts in wastewater. | Economic estimates: unknown; gypsum by-product and liquid fertilizer or $HNO_3$. | Electrical usage: unknown (will be 5-10% of total output); uses large amounts of gas phase oxidant and by-product treatment materials. |

TABLE 3.3-2. (Continued)

| | Boiler operation and/or safety | Reliability | Status of development | Adaptability to existing sources | Compatibility with other control systems |
|---|---|---|---|---|---|
| SCR Parallel Flow | $H_2$ is potential safety hazard. | Process steps well established. Should be reliable. | Both $SO_x$ and $NO_x$ removal systems have been tested on oil. $SO_2$ workup and $H_2$ generation steps not tested but are established technology. | Will need land for equipment | Compatible with particulate control systems |
| Adsorption | Possible safety hazard due to poor char distribution in beds. | Particulate plugging; numerous process steps; some corrosion resistant material in high temperature areas. | No tests on oil-fired gas. | Need land for pieces of process equipment | Not useful with FGD systems $NO_x$ removal is secondary. |
| Electron Beam Radiation | Radiation safety hazards are unknown as are those of by-product. | Few process steps; stainless steel reactor. | One oil-fired pilot plant; by-product treating method is uncertain. | Need land for pieces of process equipment. | Operate with or without FGD but capital cost is same. |
| Absorption-Reduction | No safety hazards. | Many process steps; much glass - and elastomer-lined equipment. | 3 pilot plants treating oil-fired flue gas; $NO_x$ absorption mechanism uncertain. | Much land needed for numerous pieces of process equipment. | Cannot be used in conjunction with FGD. |
| Oxidation-Absorption-Reduction | Gas phase oxidant presents serious safety hazard. | Numerous process steps and corrosion resistant material; oxidant generation system subject to periodic failure. | 6 prototype units treating oil-fired flue gas in operation. | Much land needed for numerous pieces of process equipment, oxidant generation and wastewater treatment. | Cannot be used in conjunction with FGD. |
| Oxidation-Absorption | Gas phase oxidant presents serious safety hazards. | Numerous process steps and corrosion resistant material; oxidant generation system subject to periodic failure. | One bench-scale test on oil-fired flue gas. | Much land needed for numerous pieces of process equipment, oxidant generation, and wastewater treatment. | Compatible |

### 3.3.1 Moderate Reduction Controls

One system, adsorption, was eliminated because it was not capable of achieving sufficient emission reduction. The remaining systems were rated using the selection factors and the results are presented in Table 3.3.1-1. A detailed breakdown of this evaluation is contained in Tables A3.7 and A3.8. SCR fixed packed bed was selected as the $NO_x$-only candidate system for distillate-oil-fired boilers since these have low particulate emissions. For resid-fired boilers, which have higher particulate emissions, the $NO_x$-only candidate systems are SCR parallel flow and SCR moving bed and the simultaneous $NO_x/SO_x$ candidate system is SCR parallel flow.

### 3.3.2 Stringent Reduction Controls

The results of system evaluations for stringent control levels are shown in Table 3.3.2-1. The detailed evaluation breakdown is contained in Tables A3-9 and A3-10. The candidate systems are the same as for moderate control.

### 3.3.3 Intermediate Reduction Controls

The results of system evaluations for intermediate control levels are shown in Table 3.3.3-1. A detailed breakdown of the selection factor ratings is presented in Tables A3.11 and A3.12. The candidate systems are the same as for the other two levels: $NO_x$-only, SCR fixed packed bed for distillate oil plus SCR parallel flow and SCR moving bed for resid oil; simultaneous $NO_x/SO_x$, SCR parallel flow.

## 3.4 BEST CONTROL SYSTEMS FOR GAS-FIRED BOILERS

Table 3.4-1 compares all of the FGT systems as applied to gas-fired boilers for each of the selection factors. This table was used to arrive at the point values shown on the candidate selection tables.

TABLE 3.3.1-1. CANDIDATE SYSTEMS SELECTION: OIL-FIRED BOILERS - MODERATE CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 81 | yes | Distillate oil-fired boilers |
| SCR Moving Bed | 88 | yes | Residual oil-fired boilers |
| SCR Parallel Flow | 90 | yes | Residual oil-fired boilers |
| Absorption-Oxidation | 53 | no | Low rating |
| Simultaneous $NO_x/SO_x$ | | | |
| SCR Parallel Flow | 75 | yes | Residual oil-fired boilers |
| Adsorption | NA | no | |
| Electron Beam Radiation | 47 | no | Low rating |
| Absorption-Reduction | 58 | no | Low rating |
| Oxidation-Absorption-Reduction | 59 | no | Low rating |
| Oxidation-Absorption | 52 | no | Low rating |

NA - Not applicable (see Appendix).

TABLE 3.3.2-1. CANDIDATE SYSTEMS SELECTION: OIL-FIRED BOILERS - STRINGENT CONTROLS

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 74 | yes | Distillate oil-fired boilers |
| SCR Moving Bed | 81 | yes | Residual oil-fired boilers |
| SCR Parallel Flow | 83 | yes | Residual oil-fired boilers |
| Absorption-Oxidation | NA | no | |
| Simultaneous $NO_x/SO_x$ | | | |
| SCR Parallel Flow | 71 | yes | |
| Adsorption | NA | no | |
| Electron Beam Radiation | NA | no | |
| Absorption-Reduction | NA | no | |
| Oxidation-Absorption-Reduction | 54 | no | Low rating |
| Oxidation-Absorption | 50 | no | Low rating |

NA - Not applicable (see Appendix).

TABLE 3.3.3-1. CANDIDATE SYSTEMS SELECTION: OIL-FIRED BOILERS - INTERMEDIATE CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 79 | yes | Distillate oil-fired boilers |
| SCR Moving Bed | 86 | yes | Residual oil-fired boilers |
| SCR Parallel Flow | 88 | yes | Residual oil-fired boilers |
| Absorption-Oxidation | 54 | no | Low rating |
| Simultaneous $NO_x/SO_x$ | | | |
| SCR Parallel Flow | 73 | yes | Residual oil-fired boilers |
| Adsorption | NA | no | |
| Electron Beam Radiation | 47 | no | Low rating |
| Absorption-Reduction | 58 | no | Low rating |
| Oxidation-Absorption-Reduction | 55 | no | Low rating |
| Oxidation-Absorption | 48 | no | Low rating |

NA - Not applicable (see Appendix).

## TABLE 3.4-1. COMPARISON INFORMATION OF $NO_x$-ONLY SYSTEMS FOR GAS-FIRED BOILERS

| | Performance | Operational and maintenance impacts | Preliminary environmental impacts | Preliminary economic impacts | Preliminary energy and material impacts |
|---|---|---|---|---|---|
| SCR Fixed Packed Bed | Capable of achieving >90% $NO_x$ removal; good load following capability. | Few moving parts; gas phase chemistry; simple process - good controllability; large pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | 20 MW estimates for clean gas:<br>Capital: $27/kW[1]<br>Operating: 1.2 mills/kWh | Electrical usage: unknown for gas-fired flue gas; light $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio). |
| SCR Moving Bed | Capable of achieving >90% $NO_x$ removal; adequate load following capability. | Some moving parts, solids handling - increased maintenance; gas phase chemistry; simple - controllable; low pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | Economic estimates: unknown for gas-fired plant. | Electrical usage: unknown for gas-fired flue gas; light $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio). |
| SCR Parallel Flow | Capable of achieving >90% $NO_x$ removal; good load following capability. | Few moving parts; gas phase chemistry; simple process - good controllability; moderate pressure drop. | Potential for some $NH_3$ and $NH_4HSO_4$ emissions. | Economic estimates: unknown for gas-fired plant. | Electrical usage: unknown for gas-fired flue gas; light $NH_3$ demand (1:1 $NH_3$:$NO_x$ mole ratio). |
| Absorption-Oxidation | No removal data are available; fair load following capability. | Complex process with sensitive control needs; very large pressure drop. | $NO_3^-$ salts in wastewaters. | Economic estimates: unknown. | Electrical usage: unknown; uses large amounts of liquid phase oxidant and regeneration materials. |

TABLE 3.4-1. (Continued)

| | Boiler operation and/or safety | Reliability | Status of development | Adaptability to existing sources | Compatibility with other control systems |
|---|---|---|---|---|---|
| SCR Fixed Packed Bed | No safety hazards. | Possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | Numerous commercial operations in Japan. | Some retrofit difficulty; few pieces of process equipment - little land required. | Compatible |
| SCR Moving Bed | No safety hazards. | Possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | Three commercial coke oven gas operations in Japan. | Some retrofit difficulty; few pieces of process equipment - little land required. | Compatible |
| SCR Parallel Flow | No safety hazards. | Possible $NH_4HSO_4$ scaling; simple - few process steps; little corrosion resistant material. | No commercial operations (many oil-fired; not necessary for gas-fired - no particulates) | Some retrofit difficulty; few pieces of process equipment - little land required; if space exists, catalyst can fit in duct. | Compatible |
| Absorption - Oxidation | Oxidant handling could be dangerous. | Numerous process steps and corrosion resistant material. | Pilot plants treating off gases from $HNO_3$ and steel plants. | Much land needed for numerous pieces of process equipment and wastewater treatment. | Compatible |

### 3.4.1 Moderate Reduction Controls

The first cut in FGT systems applied to gas-fired boilers eliminated one process due to insufficient emission reduction and five processes due to their removal of $SO_x$ which is not present in gas-fired flue gas. This can be seen in Table 3.4.1-1 which presents the results of the candidate selection. SCR fixed packed bed was chosen as the candidate system. SCR parallel flow and SCR moving bed were eliminated since their specialized ash handling characteristics are not required for this application. A detailed selection factor rating breakdown is contained in Table A3.13.

TABLE 3.4.1-1. CANDIDATE SYSTEMS SELECTION:
GAS-FIRED BOILERS - MODERATE CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 93 | Yes | |
| SCR Moving Bed | 91 | No | These specialized catalyst arrangements are not necessary for gas-fired sources. |
| SCR Parallel Flow | 93 | No | |
| Absorption - Oxidation | 58 | No | Low rating |

### 3.4.2 Stringent Reduction Controls

The results of system evaluations for stringent control levels are shown in Table 3.4.2-1. The detailed evaluation breakdown is contained in Table A3.14. The candidate system is SCR fixed packed bed.

TABLE 3.4.2-1. CANDIDATE SYSTEMS SELECTION:
GAS-FIRED BOILERS - STRINGENT CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 83 | Yes | |
| SCR Moving Bed | 81 | No | These specialized catalyst arrangements are not necessary for gas-fired sources. |
| SCR Parallel Flow | 83 | No | |
| Absorption - Oxidation | NA | No | |

### 3.4.3 Intermediate Reduction Controls

The results of system evaluations for intermediate control levels are shown in Table 3.4.3-1. The detailed evaluation is presented in Table A3.15. The candidate system is SCR fixed packed bed.

TABLE 3.4.3-1. CANDIDATE SYSTEMS SELECTION:
GAS-FIRED BOILERS - INTERMEDIATE CONTROL

| Control technique | Total point rating | Candidate system | Comments |
|---|---|---|---|
| $NO_x$-Only | | | |
| SCR Fixed Packed Bed | 87 | Yes | |
| SCR Moving Bed | 85 | No | These specialized catalyst arrangements are not necessary for gas-fired sources. |
| SCR Parallel Flow | 87 | No | |
| Absorption - Oxidation | 58 | No | Low rating |

## 3.5 SUMMARY

A candidate or set of candidates has now been chosen for each of the standard boilers under consideration. These are shown in Table 3.5-1. These systems will be analyzed in detail in the subsequent sections in order to determine the best overall system for $NO_x$ reduction by FGT on industrial boilers. The major performance characteristics for the candidate processes are presented in Table 3.5-2.

TABLE 3.5-1. SUMMARY OF CANDIDATE SYSTEMS: ALL LEVELS OF CONTROL

| Fuel | Candidate Systems |
|---|---|
| Coal | SCR Parallel Flow |
| Residual Oil | SCR Parallel Flow<br>SCR Moving Bed |
| Distillate Oil | SCR Fixed Packed Bed |
| Natural Gas | SCR Fixed Packed Bed |

TABLE 3.5-2. MAJOR PERFORMANCE CHARACTERISTICS OF CANDIDATE SYSTEMS

| | Collection Efficiency | Environmental Impacts | Energy Impacts | Reliability | Commercial Availability |
|---|---|---|---|---|---|
| SCR-fixed packed bed | >90% $NO_x$ reduction | Possible $NH_3$ and $NH_4HSO_4$ emissions | Coal-need ESP elec.= 1.2% of total power input | Simple, few process steps; catalyst easily plugged | Coal-bench scale; 8 oil- and numerous gas-fired commercial operations |
| SCR-parallel flow | >90% $NO_x$ reduction | Possible $NH_3$ and $NH_4HSO_4$ emissions | Coal-no ESP elec.=0.2% of total power output | Simple, few process steps; little plugging | Coal-bench scale; numerous oil-fired commercial operations |
| SCR-moving bed | >90% $NO_x$ reduction | Possible $NH_3$ and $NH_4HSO_4$ emissions | Coal-some particulate removal needed elec.<1% of total power output | Simple, few process steps; moving parts and solids handling; some plugging | Coal-bench scale; 5 oil-fired and 3 coke oven gas commercial operations |

## REFERENCES

1. Faucett, H.L., *et al.* Technical Assessment of $NO_x$ Removal Processes for Utility Applications. EPA-600/7-77/127. November 1977. pp. 210, 268, 301.

2. Rosenberg, H.S., *et al.* State-of-the-Art Review of Stack Gas Treatment Techniques for $NO_x$ Control. EPRI-Batelle. April 1976. p. A-47.

3. Faucett, H.L., *op.cit.*, pp. 233, 319.

4. *Ibid.*, p. 218.

5. *Ibid.*, p. 204.

6. *Ibid.*, p. 166.

7. *Ibid.*, p. 33

8. *Ibid.*, p. 347.

9. Rosenberg, H.S., *op.cit.*, p. A-14.

10. Faucett, H.L., *op.cit.*, pp. 280, 327.

11. Rosenberg, H.S., *op.cit.*,p. A-23.

12. Faucett, H.L., *op.cit.*,p. 242.

13. Faucett, H.L., *op.cit.*, p. 223.

14. *Ibid.*, p. 41

15. Rosenberg, H.S., *op.cit.*, p. A-139.

16. *Ibid.*, pp. A-106, A-116.

17. Faucett, H.L., *op.cit.*, pp. 48, 117, 127.

# SECTION 4

# COST ANALYSIS OF CANDIDATES FOR BEST EMISSION CONTROL SYSTEMS

## 4.1 $NO_x$-ONLY SYSTEMS

### 4.1.1 Introduction

This section considers the costs involved with applying the "best" $NO_x$ FGT systems selected in Section 3 to the standard boilers. The costs presented are based on several factors. First, typical process layouts were determined to establish the equipment requirements. Material balances are established for each case and the equipment sized. Process layouts and material balances for all nineteen cases considered in detail are presented in Appendices 3, 4 and 5 for coal, oil, and gas sources, respectively. Purchased equipment lists for each process considered are shown in Table 4.1.1-1. The equipment is selected and sized by using standard engineering techniques. Example calculations for equipment size and energy usage are presented in the Appendix 8. Energy usage for all systems consists only of electricity and steam. Other costs were based on cost factors supplied by references 1 and 2 as well as other sources.

All of the equipment listed in Table 4.1.1-1 will require some maintenance. The items requiring the most maintenance are the pump, fan motor drive, vaporizer, screen, catalyst elevator, baghouse/blower, and all associated process control elements. The catalyst has a lifetime of about one year and its regeneration is presently uncertain. Therefore, in this analysis, it is replaced annually and represents both a capital and operating cost.

TABLE 4.1.1-1. PURCHASED EQUIPMENT FOR $NO_x$ FGT SYSTEMS

| Parallel Flow SCR | Moving Bed SCR | Fixed Packed Bed SCR |
|---|---|---|
| Reactor | Reactor | Reactor |
| Catalyst | Catalyst | Catalyst |
| Fan Motor Drive | Catalyst Screen | Fan Motor Drive |
| $NH_3$ Storage Tank | Catalyst Elevator | $NH_3$ Storage Tank |
| $NH_3$ Transfer Pump | Baghouse/Blower | $NH_3$ Transfer Pump |
| $NH_3$ Vaporizer | Fan Motor Drive | $NH_3$ Vaporizer |
|  | $NH_3$ Storage Tank |  |
|  | $NH_3$ Transfer Pump |  |
|  | $NH_3$ Vaporizer |  |

The cost bases can be separated into several areas. Costs of materials associated with all of the processes evaluated are presented in Table 4.1.1-2. Sources of the costs are also shown. Several costs were determined by multiplying a factor times another cost. This is common with this type of economic analysis and the cost factors used are shown in Table 4.1.1-3. Direct costs were determined on a full year basis and then multiplied by the boiler load factor to determine the annual direct costs. Load factors for the standard boilers are shown in Table 4.1.1-4. The capital recovery factor was calculated from the formula:

$$\frac{i(1+i)^n}{(1+i)^n-1}$$

For i = 0.10 (interest) and n = 15 (years) the capital recovery factor is 0.13147.

The costs of each equipment item was determined using a variety of cost references shown in Table 4.1.1-5. Installation costs were provided in the references. As with the annual costs, some of the capital costs were determined by multiplying a factor times another cost. A list of the factors used in the capital cost estimates is contained in Table 4.1.6.

TABLE 4.1.1-2. ANNUAL COST PARAMETERS USED IN COST ANALYSIS

| Item | Cost Used | Reference |
|---|---|---|
| Direct Labor, $/manhour | 12.02 | 1 |
| Maintenance Labor, $/manhour | 14.63 | 1 |
| Electricity, mills/kWh | 25.8 | 1 |
| Ammonia, $/ton delivered | 130 | 1 |
| Steam, $/1000 lb | 3.50 | 2 |
| Catalyst, $/ft$^3$ | | |
| Parallel Flow | 212 | 3 |
| Moving Bed | 282 | 4 |
| Fixed Packed Bed | 282 | 4 |

TABLE 4.1.1-3. ANNUAL COST FACTORS

| Item | Amount | Reference |
|---|---|---|
| Maintenance Materials | 3% of turnkey costs | 5 |
| Payroll Overhead | 30% of direct labor | 1 |
| Plant Overhead | 26% of labor, parts & maintenance | 1 |
| General and Administrative Expenses (G&A), Taxes & Insurance | 4% of total turnkey costs | 1 |
| Capital Recovery Factor (10% interest rate) | 13.147% of total turnkey costs | 1 |

TABLE 4.1.1-4. LOAD FACTORS FOR THE STANDARD BOILERS

| Fuel | Load Factor |
|---|---|
| Coal | 0.60 |
| Residual Oil | 0.55 |
| Distillate Oil and Natural Gas | 0.45 |

TABLE 4.1.1-5. SOURCES OF COSTS FOR SPECIFIC EQUIPMENT ITEMS

| Equipment Item | Reference |
|---|---|
| Reactor | 6 |
| Catalyst | 3,4 |
| Fan Motor Drive (Incremental) | 7,8 |
| $NH_3$ Storage Tank | 9 |
| $NH_3$ Transfer Pump | 10 |
| $NH_3$ Vaporizer | 11 |
| Vibrating Screen | 12 |
| Catalyst Elevator | 13,8 |
| Baghouse | 14 |

TABLE 4.1.1-6. CAPITAL COST FACTORS

| Item | Amount | Reference |
|---|---|---|
| Engineering | 10% of installed cost of largest $NO_x$ removal system considered (pulverized coal boiler; stringent level of control) | 1 |
| Construction and Field Expense | 10% of installed cost | 1 |
| Contractor Fee | 10% of installed cost | 1 |
| Start-up | 2% of installed cost | 1 |
| Performance Tests | $2000 | 1 |
| Contingency | Coal: 20% of total direct and indirect costs<br>Oil and Gas: 15% of total direct and indirect costs | 1 |
| Working Capital | 25% of total direct operating costs | 1 |

Capital costs were escalated to June 1978 costs using standard cost indices. For example, costs in Guthrie[6] are based on June 1970 costs. Cost indices for this year and June 1978 for various types of equipment are shown in Table 4.1.1-7. To obtain the mid 1978 costs the costs given in Guthrie are multiplied by the escalation index.

TABLE 4.1.1-7. CHEMICAL ENGINEERING COST INDICES[15]

| Item | June 1970 Index | June 1978 Index | Escalation Index (E.I.) (1978/1970) |
|---|---|---|---|
| Fabricated Equipment | 124.0 | 237.4 | 1.91 |
| Process Machinery | 122.7 | 226.6 | 1.85 |
| Pipes, Valves & Fittings | 133.0 | 268.4 | 2.02 |
| Process Instruments | 132.0 | 214.8 | 1.63 |
| Pumps & Compressors | 124.1 | 258.2 | 2.08 |
| Electrical Equipment | 98.9 | 167.9 | 1.70 |
| Miscellaneous | 118.5 | 250.1 | 2.11 |
| Construction Labor | 134.8 | 184.3 | 1.37 |

The labor requirements were determined from the basis for an economic analysis performed by a process vendor which indicated a requirement of one person/shift/day per reactor.[16] Equipment life was estimated at 15 years based on the average lifetime of chemical processing equipment.[17] Capital costs were annualized over a 15 year period to give constant annual costs for the life of the boiler.

The capital and operating costs were collected and presented in a consistent set of tables and an annualized cost was calculated. These comprehensive tables are contained in separate appendices and discussed in the subsequent subsections.

Costs for modified or reconstructed facilities will most likely be slightly higher than those for new facilities. This is due to the fact the major cost items--*i.e.* the fan motor drive, reactor plus catalyst, and $NH_3$ storage tanks--are the same for both applications. There may be some increased costs where additional ductwork, boiler modification or flue gas heating is necessary and these factors are highly site specific. The cost of a retrofit will have to be determined for each application since it is dependent on site specific factors.

The cases considered include only one type of coal, low sulfur western. Other coal types are not considered since process costs do not vary significantly with coal type. Two of the most significant cost items for FGT systems are the reactor plus catalyst and the fan motor drive. These equipment items are sized and costed based on flue gas flow rate which does not vary significantly with coal type. Since including all three coal types would not provide additional information, only low sulfur western was considered. Since all catalysts considered in this study are assumed to be resistant to $SO_x$ poisoning, low sulfur coal was chosen since it had the highest $NO_x$ emissions and flue gas flow rate. Therefore, use of this coal provides a "worst case" analysis.

SIP control levels are not considered since in many cases no control is required. On cases that require some control the level can be easily attained through use of combustion modifications. In no instance is FGT required to meet the average SIP levels, except possibly in California. Los Angeles hourly maximum $NO_x$ concentration occassionally exceeds the state standard by a factor of 3.[18] Because of the topographic characteristics of the area and its high concentration of mobile sources (automobiles which also have mileage requirements to meet), strict legislation for stationary sources has been proposed that would require $NO_x$ FGT on boilers.

### 4.1.2 Control Costs for Coal-Fired Boilers

Equipment costs are determined from equipment sizing calculations which are in turn determined from material balances. Material balances for coal-fired boilers are contained in Appendix 3. These and the factors discussed in Section 4.1.1 were used to compute the various cost values. The cost components are broken down into individual capital and operating costs in Appendices 6 and 7, respectively.

The annualized costs for each of the standard boilers are summarized in Tables 4.1.2-1 through 4.1.2-4. The costs are also presented as a percent of cost of the uncontrolled boiler. These data are also plotted in Figures 4.1.2-1 through 4.1.2-4 to show the sensitivity of the process costs to control level. The slight nonlinearities are a result of the cost of catalyst which increases while several equipment costs and labor costs are constant for all control levels.

The cost effectiveness of the various applications can be assessed by dividing the annual cost by the annual $NO_x$ removal. The results of this calculation are presented in Table 4.1.2-5. As can be seen, the effectiveness of the parallel flow system on the largest boiler indicates an optimum at 70% while the smallest boiler exhibits an optimum at 90%. There are three primary cost components that determine these results: equipment costs per unit size, catalyst costs, and labor costs. On the smaller boilers the equipment required is obviously smaller and its costs per unit size is greater due to the lack of economy of scale. This is shown directly by the improving cost effectiveness with boiler size (less $ required per kg $NO_x$ removed). Now, on the smaller boilers, the catalyst costs are not as dominant as the labor costs. This is due to less catalyst required by the smaller boilers, yet the operating and maintenance labor requirements for the $NO_x$ systems on smaller boilers are comparable to those of larger boilers (at least within the size range of the standard boilers). What this means is that on a small $NO_x$ system where maintenance and operation

TABLE 4.1.2-1. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Annual costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Standard boilers | | | Type and level of control† | Control efficiency (%) | | | % increase in costs over uncontrolled boiler |
| Heat input | | Type | | | | | |
| MW | (MBtu/hr) | | | | $/J/S | ($/MBtu/hr) | |
| 8.8<br>Low Sulfur Western Coal | (30) | Package Watertube Underfeed Stoker | PF SCR Intermediate | 80 | 0.0134 | (3920) | 10.7 |

*Based only on Annual Costs

†PF = Parallel Flow SCR

TABLE 4.1.2-2. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Annual costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Heat input | | | | | | | |
| MW | (MBtu/hr) | Type | Type and level of control† | Control efficiency (%) | \$/J/S | (\$/MBtu/hr) | % increase in costs over uncontrolled boiler |
| 22 | (75) | Package Watertube Chaingrate | PF SCR Stringent | 90 | 0.00882 | (2620) | 9.1 |
| Low Sulfur Western Coal | | | PF SCR Intermediate | 80 | 0.00769 | (2270) | 7.9 |
| | | | PF SCR Moderate | 70 | 0.00687 | (2030) | 7.1 |

*Based only on Annual Costs.
†PF = Parallel Flow SCR

## TABLE 4.1.2-3. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Annual costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Heat input | | | | | | | % increase in costs over uncontrolled boiler |
| MW | (MBtu/hr) | Type | Type and level of control† | Control efficiency (%) | \$/J/S | (\$/MBtu/hr) | |
| 44<br>Low Sulfur Western Coal | (150) | Field-Erected Watertube Spreader Stoker | PF SCR Intermediate | 80 | 0.00567 | (1680) | 7.2 |

*Based only on Annual Costs

†PF = Parallel Flow SCR

## TABLE 4.1.2-4. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Annual costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Heat input | | | | | | | % increase in costs over uncontrolled boiler |
| MW | (MBtu/hr) | Type | Type and level of control† | Control efficiency (%) | \$/J/S | (\$/MBtu/hr) | |
| 58.6<br>Low Sulfur Western Coal | (200) | Field-Erected Watertube | PF SCR Stringent | 90 | 0.00599 | (1760) | 7.9 |
| | | Pulverized Coal | PF SCR Moderate | 70 | 0.00433 | (1270) | 5.7 |

*Based only on Annual Costs

†PF = Parallel Flow SCR

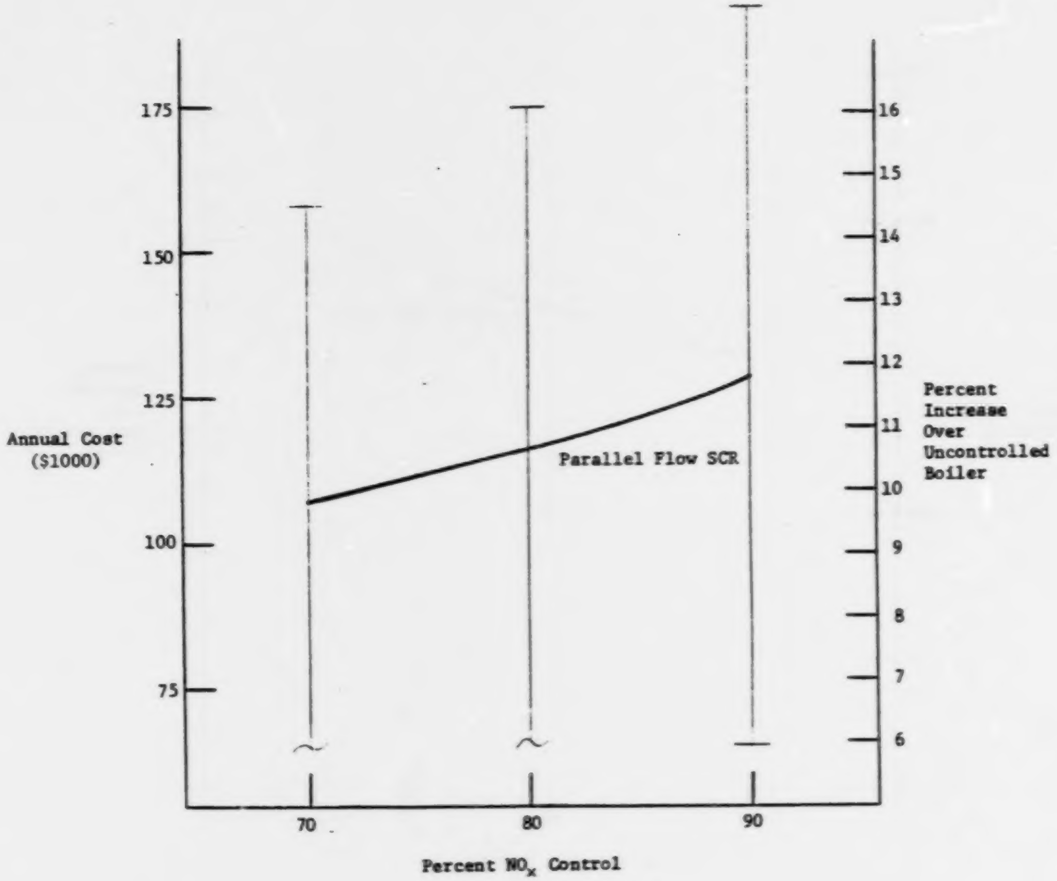

Figure 4.1.2-1. Annual cost of $NO_x$ control systems applied to underfeed stoker standard boiler.

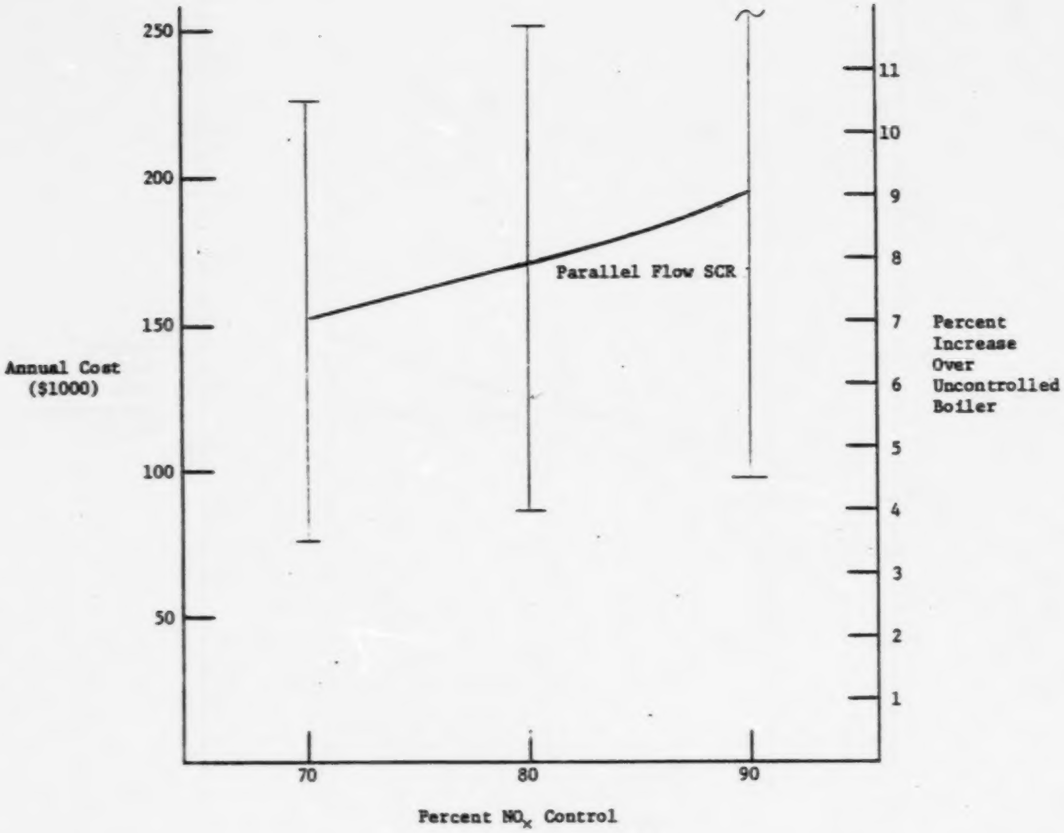

Figure 4.1.2-2. Annual cost of $NO_x$ control systems applied to chaingrate standard boiler.

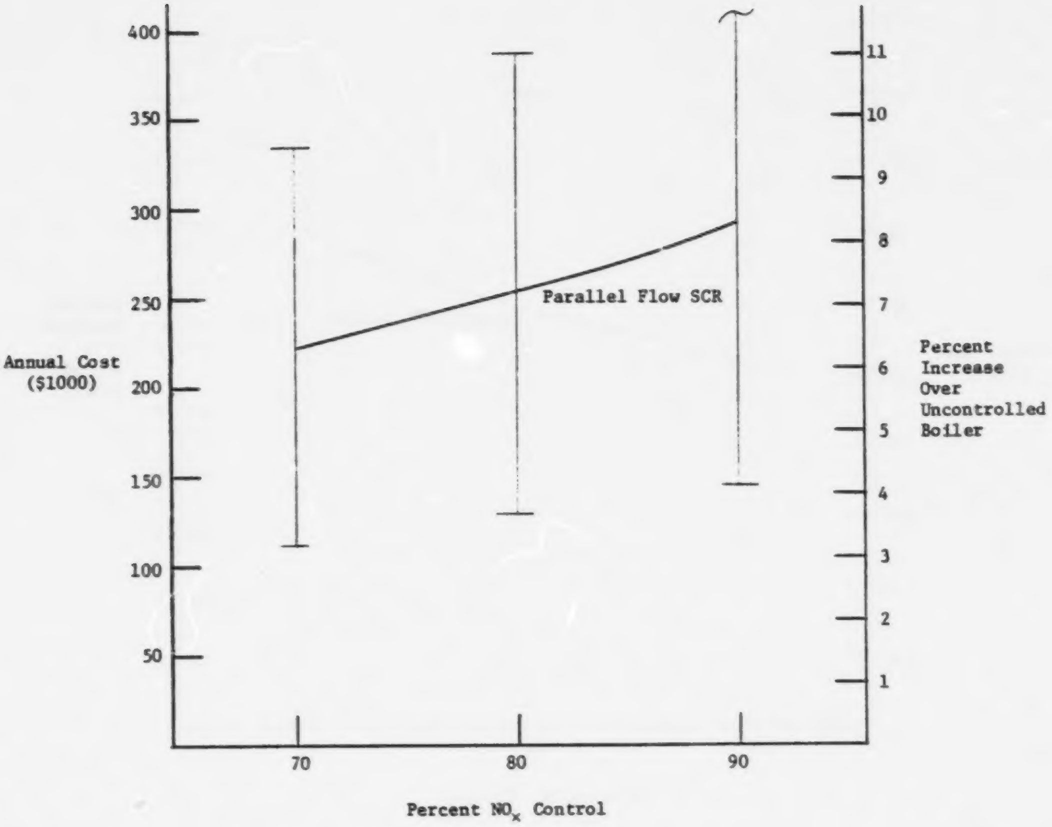

Figure 4.1.2-3. Annual cost of $NO_x$ control systems applied to spreader stoker standard boiler.

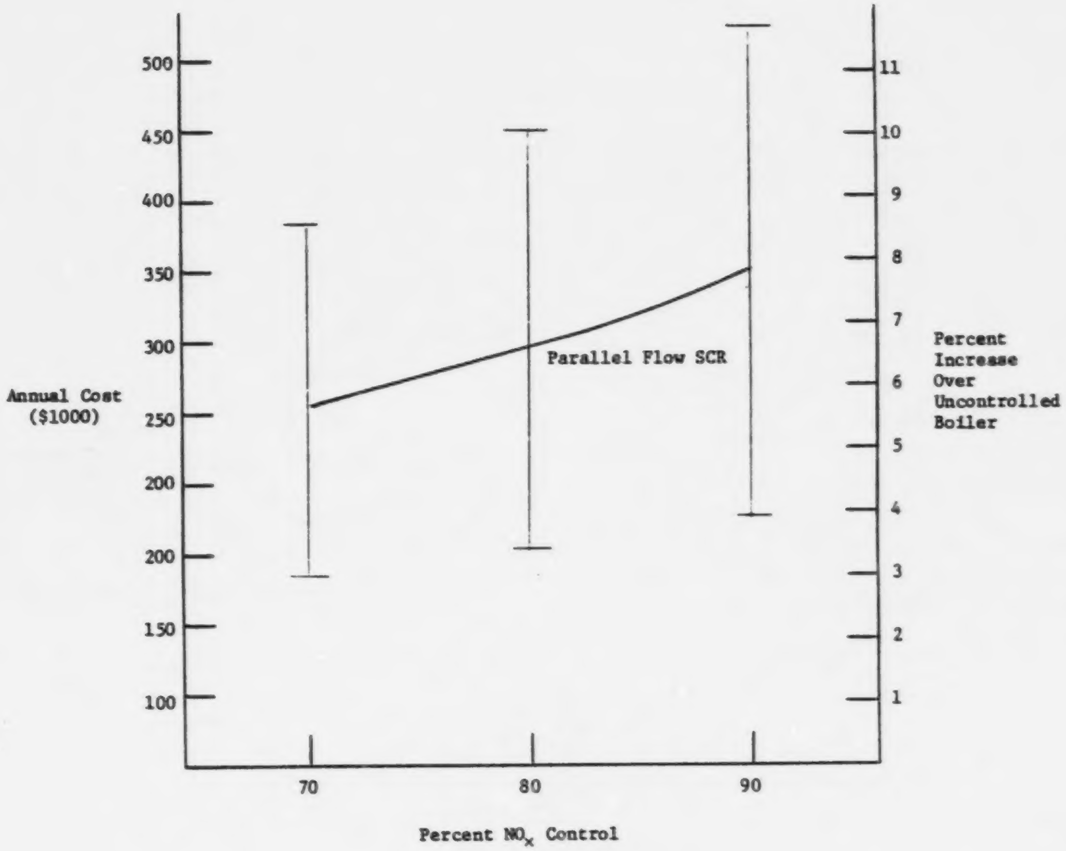

Figure 4.1.2-4. Annual cost of $NO_x$ control systems applied to pulverized coal standard boiler.

TABLE 4.1.2-5. COST EFFECTIVENESS OF $NO_x$ FGT ($/kg $NO_x$ removed)

| Boiler type | Control type | Percent $NO_x$ control<br>90 | <br>80 | <br>70 |
|---|---|---|---|---|
| Underfeed | PF | 2.57 | 2.64 | 2.75 |
| Chaingrate | PF | 1.56 | 1.53 | 1.56 |
| Spreader Stoker | PF | 1.16 | 1.13 | 1.14 |
| Pulverized Coal | PF | 0.874 | 0.836 | 0.813 |

personnel are going to be needed regardless, the $NO_x$ system might as well be a little larger to remove additional $NO_x$. This trend is just the opposite for larger boilers where catalyst costs become dominant. It requires larger amounts of expensive catalyst to remove the additional $NO_x$ and thus increases the cost substantially and decreases the cost effectiveness of the system at higher removal levels. The data presented in Table 4.1.2-5 is plotted in Figure 4.1.2-5.

The cost of applying $NO_x$ FGT to modified or reconstructed facilities is likely to be higher than the cost for applications to new facilities. All of the equipment for new installations will be necessary for retrofit installations, however, additional equipment may also be necessary. Specific costs for retrofit applications were not calculated here, but can be estimated. In a study for the Japanese Environment Agency, five process vendors prepared economic analyses for three applications:[19]

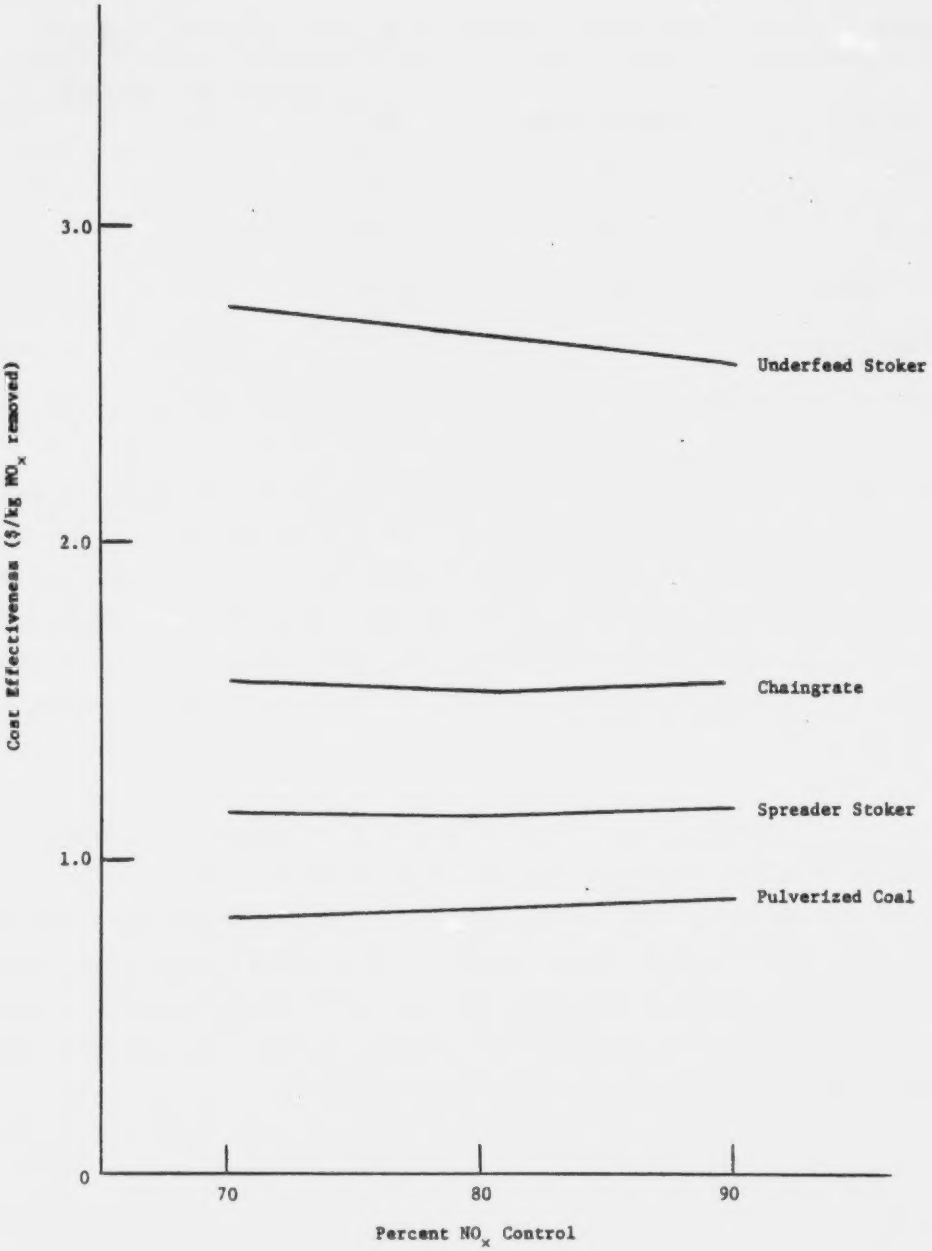

Figure 4.1.2-5. Cost effectiveness of parallel flow SCR $NO_x$ control systems applied to the coal-fired standard boilers.

1) new boiler,

2) retrofit for gas taken upstream of the air preheater requiring additional ductwork and a fan, and

3) retrofit for gas taken downstream of the ESP including gas/gas heat exchanger, heater, fan.

The relative costs for each system treating 40,000 $Nm^3$/hr of flue gas is shown in Table 4.1.2-6.

TABLE 4.1.2-6. RELATIVE COSTS OF RETROFIT SCR SYSTEMS

| System | Relative annualized cost |
|---|---|
| 1 | 1.00 |
| 2 | 1.23 |
| 3 | 2.20 |

These results indicate that SCR applications to modified or reconstructed facilities can cost from 25 to 120 percent more than applications to new boilers.

### 4.1.3 Costs To Control Oil-Fired Boilers

The cost calculations presented in this section are based on material balances performed for each case considered. The material balances are presented in Appendix 4. These are used to size the equipment which is subsequently costed. The costing techniques are described in principle in Section 4.1.1.

The cost components are broken down into individual capital and operating costs in Appendices 6 and 7. The annualized costs are summarized in Tables 4.1.3-1 and 4.1.3-2. These tables show the annual cost as a percentage of the uncontrolled boiler cost. These values are plotted as a function of control level in Figures 4.1.3-1 and 4.1.3-4.

The parallel flow system shows slightly more sensitivity to control level. This is most likely due to the catalyst which is the most significant cost component. The parallel flow catalyst is about as expensive as the moving bed catalyst per cubic meter, but has a lower space velocity. This causes the parallel flow systems to have a higher catalyst cost component. The nonlinearity is due to this fact combined with the fact that the cost/unit of equipment increases as size (*i.e.*, control level) decreases.

The cost effectiveness is also determined in Table 4.1.3-3 where the cost per kg of $NO_x$ removed is presented. The cost for the distillate oil-fired boiler is very high due primarily to poor economy of scale since the boiler is small. These costs are plotted in Figures 4.1.3-5 and 4.1.3-6.

The cost differences between the two systems applied to the residual oil-fired boilers are not significant within the accuracy of this cost estimate (±50 percent). The table indicates that the cost effectiveness of the moving bed system increases as removal level increases. This seems to be due to the effect of a greater economy of scale with the larger systems. The reactor is smaller than the parallel flow so the catalyst cost is not as dominant a cost component as the labor cost component. There are several different types of parallel flow type reactors. Some of them consume more energy and cost more than the moving bed reactors, as described here. However, reactors using thin-wall honeycomb or plate catalysts developed recently in Japan are reported to require less energy and cost less than moving bed reactors, and have been used for virtually all of the new SCR plants for dirty or semi-dirty gases.

TABLE 4.1.3-1. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR OIL-FIRED BOILERS

| System | | | | | Annual Costs | | Impacts* % increase in costs over uncontrolled boiler |
|---|---|---|---|---|---|---|---|
| Standard boilers | | | Type and level of control† | Control efficiency (%) | | | |
| Heat input | | | | | | | |
| $MW_t$ | (MBtu/hr) | Type | | | $/J/S | ($/MBtu/hr) | |
| 4.4 | (15) | Package Firetube Scotch | FPB SCR Stringent | 90 | 0.0154 | (4500) | 12.1 |
| Distillate Oil | | | FPB SCR Moderate | 70 | 0.0145 | (4240) | 11.4 |
| 44 | (150) | Package Watertube | FPB SCR Stringent | 90 | 0.0040 | (1170) | 7.5 |
| | | | FPB SCR Moderate | 70 | 0.0031 | (915) | 5.6 |

*Based only on Annual Costs

†FPB = Fixed Packed Bed

TABLE 4.1.3-2. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR OIL-FIRED BOILERS

| System | | | | | Annual costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Standard boilers | | | Type and level of control† | Control efficiency (%) | | | % increase in costs over uncontrolled boiler |
| Heat input | | Type | | | | | |
| $MW_t$ | (MBtu/hr) | | | | $/J/S | ($/MBtu/hr) | |
| 8.8 | 30 | Package Watertube | PF SCR Stringent | 90 | 0.0123 | (3600) | 14 |
| Residual Fuel Oil | | | PF SCR Moderate | 70 | 0.0110 | (3200) | 12 |
| | | | MB SCR Stringent | 90 | 0.0148 | (4330) | 16 |
| | | | MB SCR Moderate | 70 | 0.0137 | (4010) | 15 |
| 44 | (150) | Package Watertube | PF SCR Stringent | 90 | 0.00502 | (1490) | 7.0 |
| Residual Fuel Oil | | | PF SCR Moderate | 70 | 0.00408 | (1210) | 5.7 |
| | | | MB SCR Stringent | 90 | 0.00457 | (1360) | 6.4 |
| | | | MB SCR Moderate | 70 | 0.00377 | (1120) | 5.3 |

* Based only on Annual Costs

†PF = Parallel Flow
MB = Moving Bed

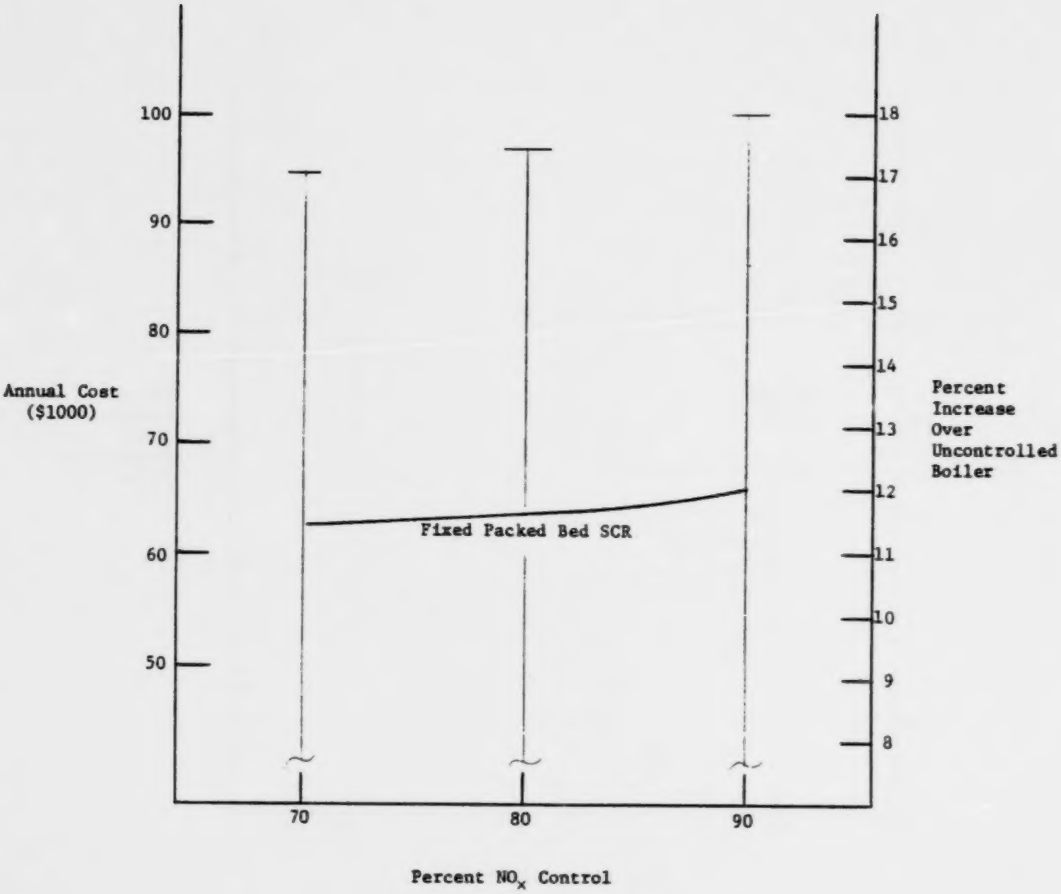

Figure 4.1.3-1. Annual cost of $NO_x$ control system applied to 4.4 MW distillate oil-fired standard boiler.

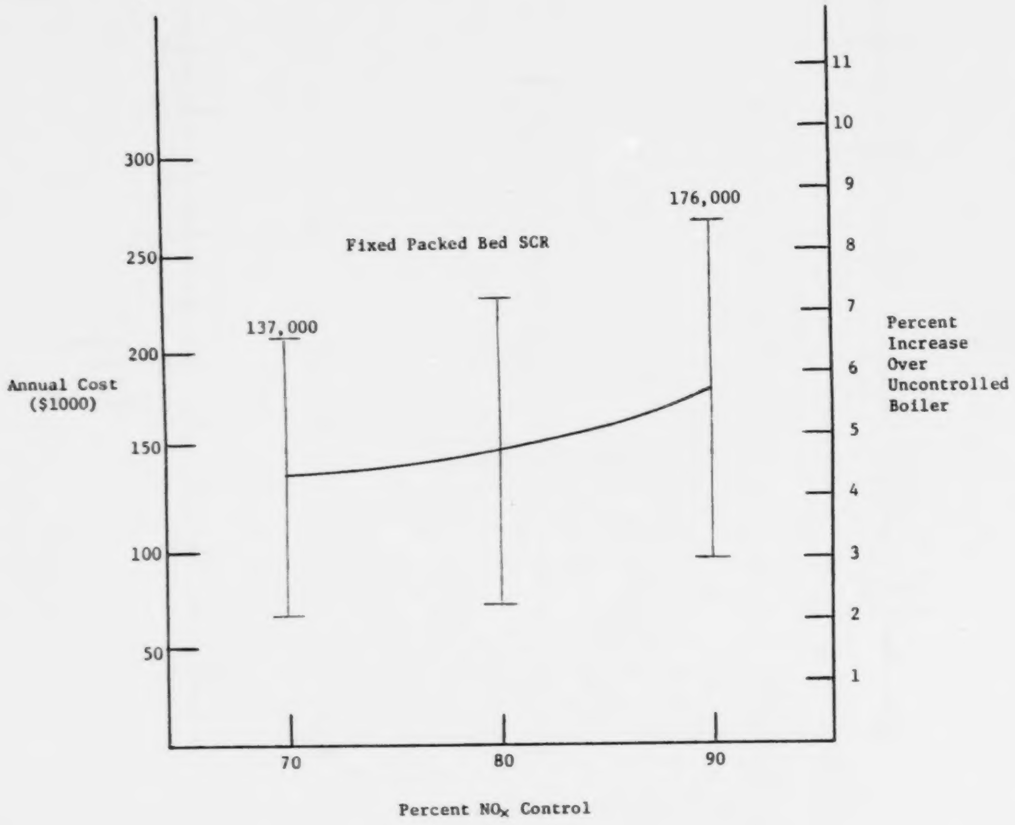

Figure 4.1.3-2. Annual cost of NO$_x$ control system applied to 44 MW distillate oil-fired standard boiler.

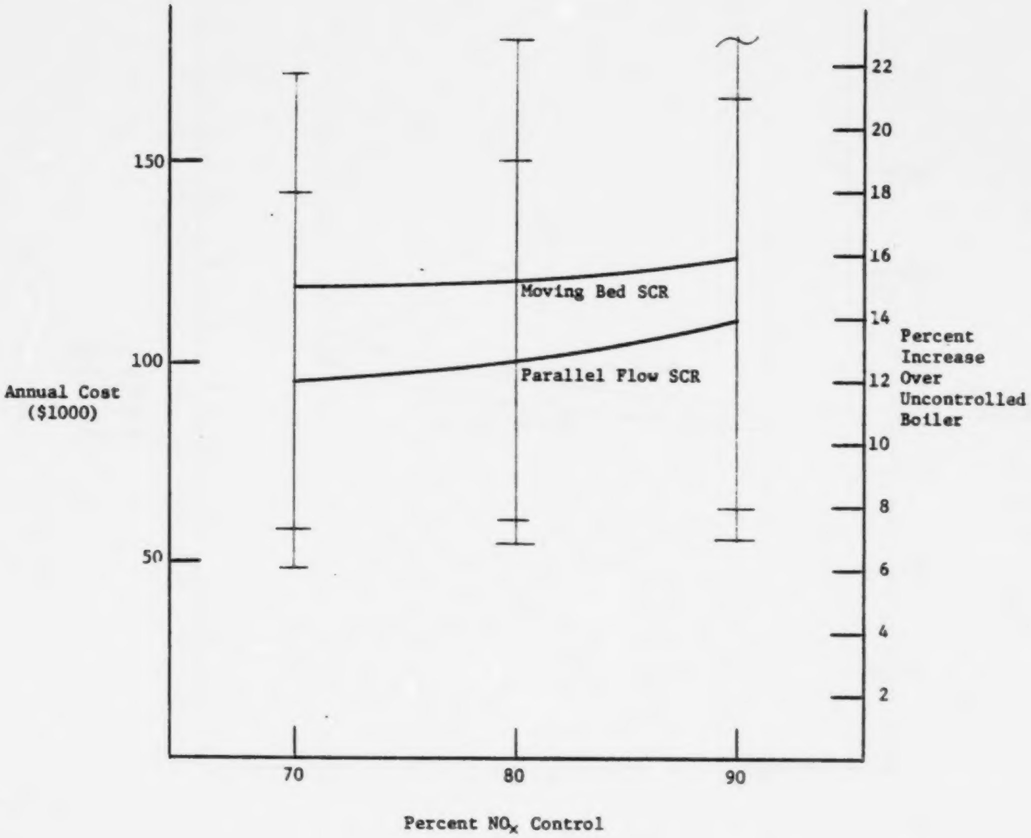

Figure 4.1.3-3. Annual cost of $NO_x$ control systems applied to 8.8 MW residual oil-fired standard boiler.

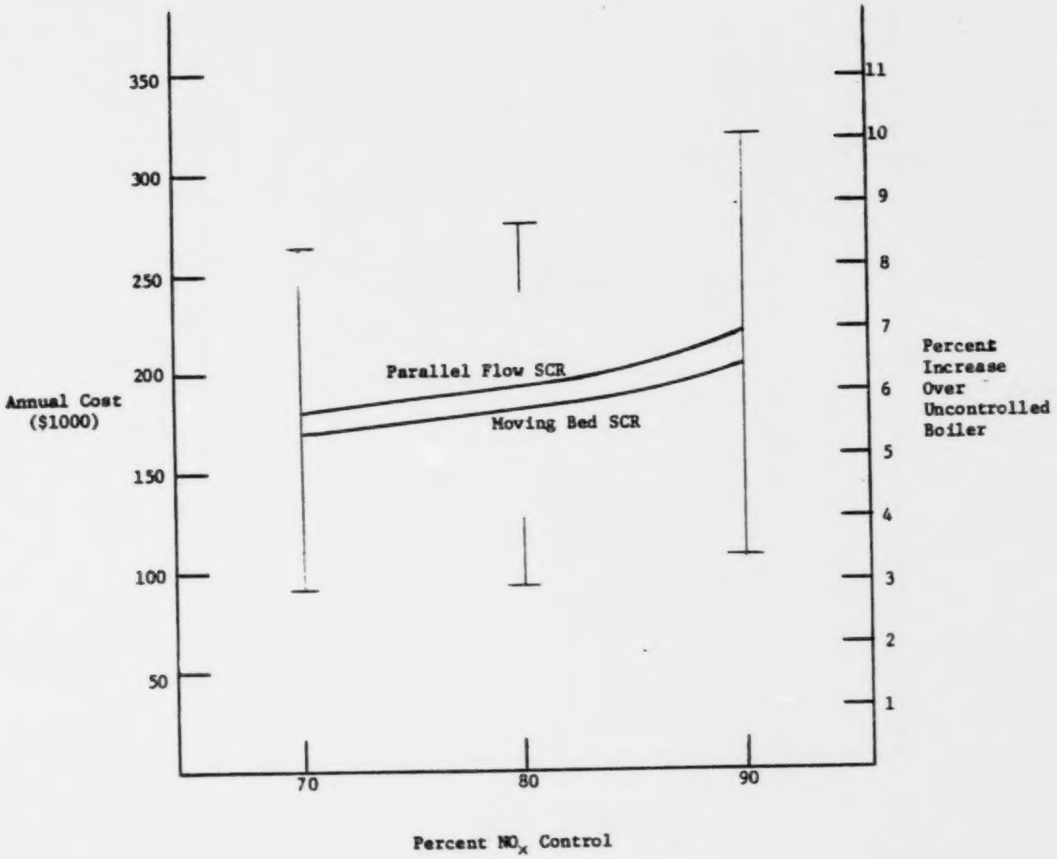

Figure 4.1.3-4. Annual cost of NO$_x$ control systems applied to 44 MW residual oil-fired standard boiler.

TABLE 4.1.3-3. COST EFFECTIVENESS OF NOx FGT

| Fuel | Boiler size, $MW_t$ | Control type* | Percent NOx control ($/kg NOx removed) | | |
|---|---|---|---|---|---|
| | | | 90 | 80 | 70 |
| Distillate Oil | 4.4 | FPB | 17.6 | 19.0 | 21.4 |
| Distillate Oil | 44 | FPB | 3.8 | 3.8 | 3.8 |
| Residual Oil | 8.8 | PF | 5.7 | 6.0 | 6.6 |
| | | MB | 6.9 | 7.3 | 8.2 |
| Residual Oil | 44 | PF | 1.89 | 1.85 | 1.97 |
| | | MB | 1.72 | 1.75 | 1.84 |

* FPB = Fixed Packed Bed SCR
PF = Parallel Flow SCR
MB = Moving Bed SCR

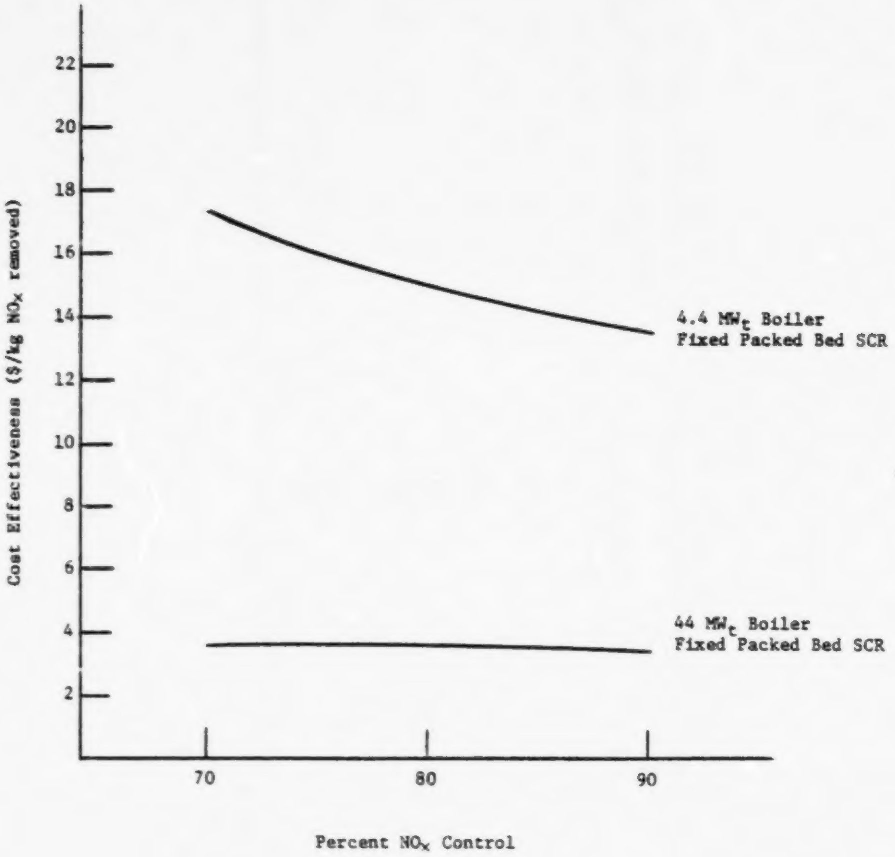

Figure 4.1.3-5. Cost effectiveness of FGT systems applied to distillate oil-fired boilers.

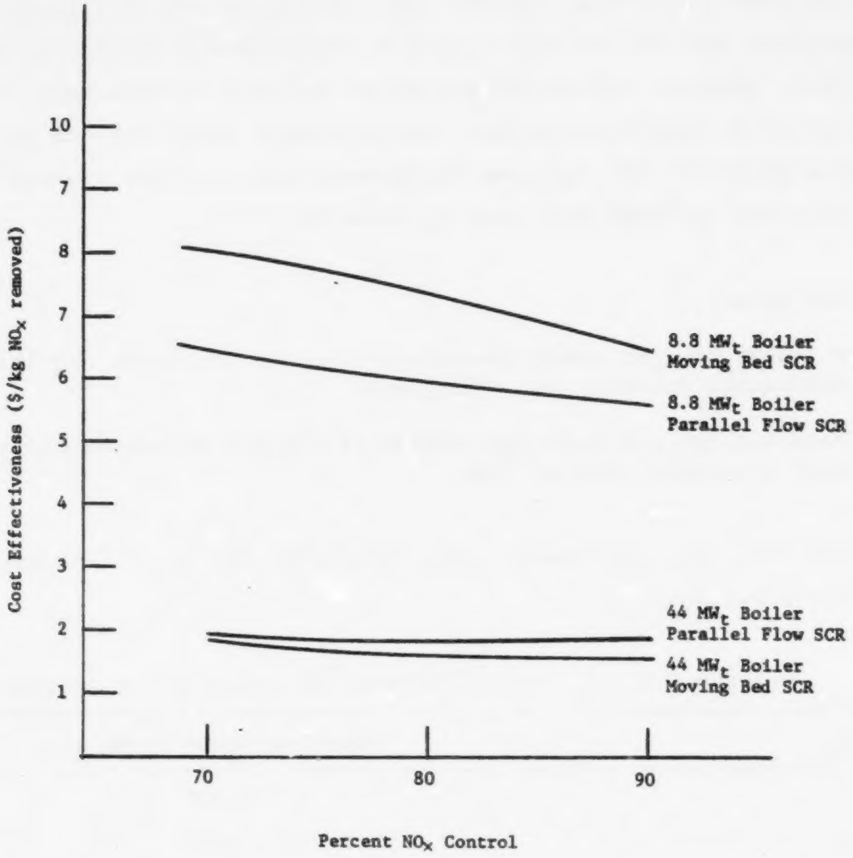

Figure 4.1.3-6. Cost effectiveness of FGT systems applied to residual oil-fired boilers.

The overall conclusion is the systems applied to the residual oil-fired boilers to not appear to be cost sensitive with respect to control level. The small distillate oil-fired boiler appears to be very sensitive since higher control can be achieved with only slightly higher annualized costs.

The cost of applying $NO_x$ FGT to modified or reconstructed facilities is likely to be higher than the cost for applications to new facilities. All of the equipment for new installations will be necessary for retrofit installations, however, additional equipment may also be necessary. Specific costs for retrofit applications were not calculated here, but can be estimated. In a study for the Japanese Environment Agency, five process vendors prepared economic analyses for three applications:[19]

1) new boiler,
2) retrofit for gas taken upstream of the air preheater requiring additional ductwork and a fan, and
3) retrofit for gas taken downstream of the ESP including gas/gas heat exchanger, heater, fan.

The relative costs for each system treating 40,000 Nm /hr of flue gas are shown in Table 4.1.3-4.

TABLE 4.1.3-4. RELATIVE COSTS OF RETROFIT SCR SYSTEMS

| System | Relative annualized cost |
|---|---|
| 1 | 1.00 |
| 2 | 1.23 |
| 3 | 2.20 |

These results indicate that SCR applications to modified or reconstructed facilities can cost from 25 to 120 percent more than applications to new boilers.

### 4.1.4 Control Costs for Natural Gas-Fired Boilers

This section presents cost calculations for a FGT system applied to the natural gas-fired standard boiler. The calculations are based on material balances contained in Appendix 5. The costing techniques have been described in Section 4.1.2.

The cost components are broken down into individual capital and operating costs in Appendices 6 and 7. Both total annualized costs and costs as a percentage of the uncontrolled boiler cost are shown in Table 4.1.4-1. The data presented are also plotted in Figures 4.1.4-1 and 4.1.4-2 to show the sensitivity to control level. The costs are fairly linear with control level indicating only a slight sensitivity.

The cost effectiveness of $NO_x$ control on natural gas-fired boilers is determined in Table 4.1.4-2 and these values are plotted in Figure 4.1.4-3. The smaller systems are not as cost effective as the larger systems since the cost per unit size of equipment is less for larger systems. The cost effectiveness of the small system is more sensitive to control level due to the influence of labor costs, which are constant for all control levels.

The cost of applying $NO_x$ FGT to modified or reconstructed facilities is likely to be higher than the cost for applications to new facilities. All of the equipment for new installations will be necessary for retrofit installations, however, additional equipment may also be necessary. Specific costs for retrofit applications were not calculated here, but can be estimated. In a study for the Japanese Environment Agency, five process vendors prepared economic analyses for three spplications:[19]

1) new boiler,
2) retrofit for gas taken upstream of the air preheater requiring additional ductwork and a fan, and

TABLE 4.1.4-1. COSTS OF $NO_x$ FGT CONTROL TECHNIQUES FOR NATURAL GAS-FIRED BOILERS

| System | | | | | Annual costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Standard boilers | | | Type and level of control† | Control efficiency (%) | | | % increase in costs over uncontrolled boiler |
| Heat input | | Type | | | | | |
| $MW_t$ | (MBtu/hr) | | | | $/J/S | ($/MBtu/hr) | |
| 4.4 Natural Gas | (15) | Package Firetube Scotch | FPB SCR Stringent | 90 | 0.0154 | (4510) | 13.6 |
| | | | FPB SCR Moderate | 70 | 0.0146 | (4290) | 13.0 |
| 44 Natural Gas | (150) | Package Watertube | FPP SCR Stringent | 90 | 0.0040 | (1160) | 7.5 |
| | | | FPB SCR Moderate | 70 | 0.0029 | (863) | 5.6 |

*Based only on Actual Costs
†FPB = Fixed Packed Bed

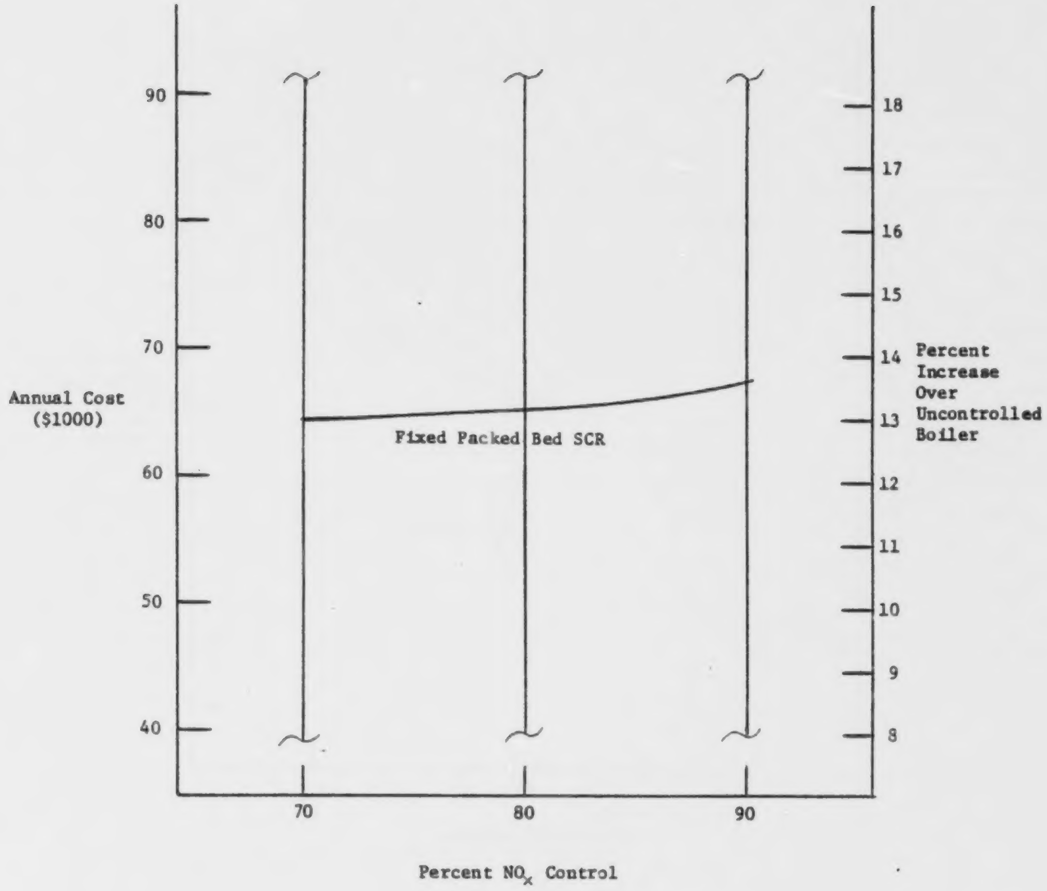

Figure 4.1.4-1. Annual cost of $NO_x$ control system applied to 4.4 MW natural gas-fired standard boiler.

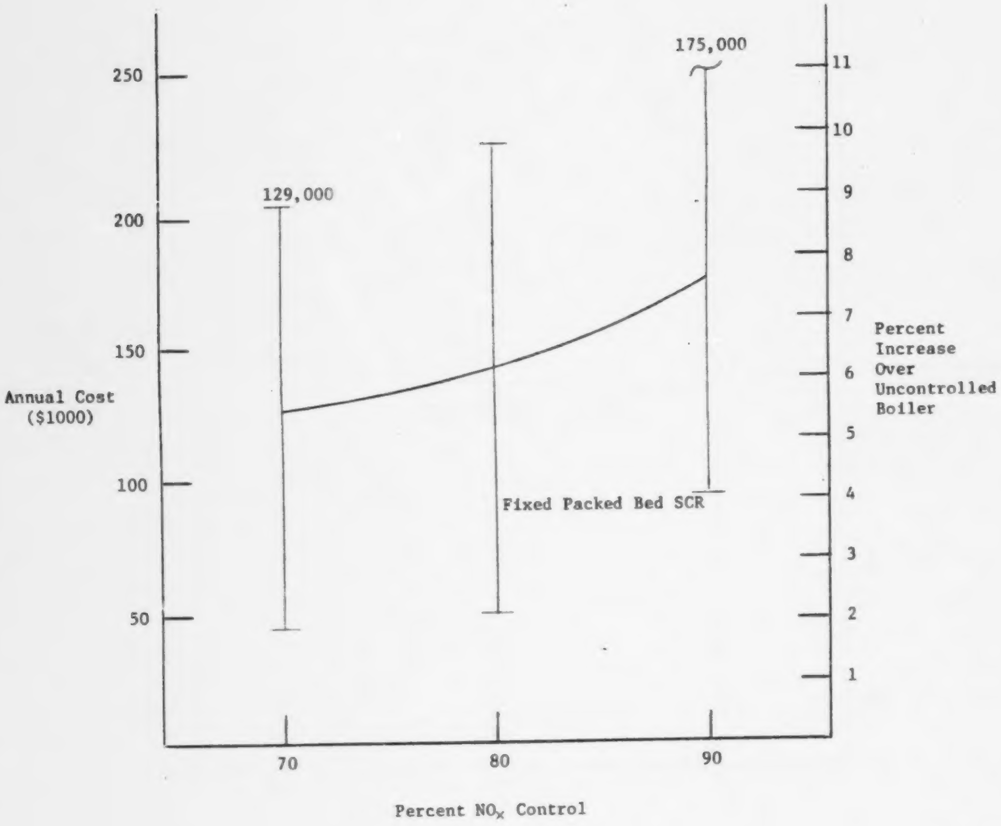

Figure 4.1.4-2. Annual cost of $NO_x$ control system applied to 44 MW natural gas-fired standard boiler.

TABLE 4.1.4-2. COST EFFECTIVENESS OF $NO_x$ FGT

| | | | $/kg $NO_x$ removed | | |
|---|---|---|---|---|---|
| | | | Percent $NO_x$ control | | |
| Fuel | Boiler size, $MW_t$ | Control type | 90 | 80 | 70 |
| Natural Gas | 4.4 | Fixed Packed Bed | 16.0 | 17.5 | 19.7 |
| Natural Gas | 44 | Fixed Packed Bed | 3.4 | 3.3 | 3.2 |

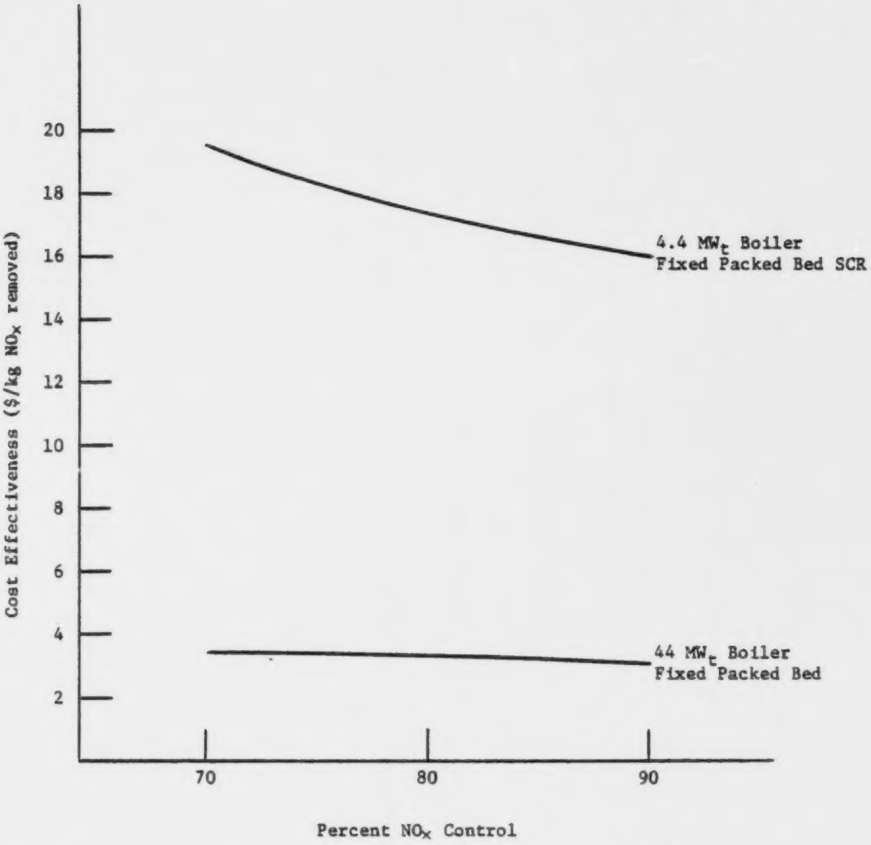

Figure 4.1.4-3. Cost effectiveness of FGT systems applied to natural gas-fired boilers.

3) retrofit for gas taken downstream of the air preheater including gas/gas heat exchanger, heater, fan.

The relative costs for each system treating 40,000 $Nm^3$/hr of flue gas is shown in Table 4.1.4-3.

TABLE 4.1.4-3. RELATIVE COSTS OF RETROFIT SCR SYSTEMS

| System | Relative annualized cost |
|---|---|
| 1 | 1.00 |
| 2 | 1.23 |
| 3 | 2.20 |

These results indicate that SCR applications to modified or reconstructed facilities can cost from 25 to 120 percent more than applications to new boilers.

### 4.1.5 Summary

In all cases the catalyst cost is a significant capital cost. Other significant capital cost components are labor, fan motor drive, and $NH_3$ storage tanks. The most significant operating cost component in all cases was labor. The smaller systems are more significantly affected by this fact than are the larger systems. As a result, the costs for small systems are high, not only because they lack economy of scale, but due to labor considerations as well. As a result the size of the unit has a greater effect on costs than does control level.

This dramatic effect is most readily observed in the cost effectiveness numbers. The systems exhibited an order of magnitude larger cost/kg of $NO_x$ when applied to the smallest systems.

## 4.2 $NO_x/SO_x$ SYSTEM

### 4.2.1 Introduction

This section considers the costs of applying the $NO_x/SO_x$ system selected in Section 3 to two coal-fired and one oil-fired boilers. The costing techniques are the same as used with the $NO_x$-only processes and will not be repeated here. The equipment items are more numerous due to the higher number of process operations associated with the process. These items are listed in Table 4.2.1-1.

TABLE 4.2.1-1. PURCHASED EQUIPMENT FOR $NO_x$ FGT SYSTEMS

| $NO_x/SO_x$ Parallel Passage |
|---|
| Reactors (2) |
| Catalyst |
| Fan Motor Drive |
| $NH_3$ Storage Tank |
| $NH_3$ Transfer Pump |
| $NH_3$ Vaporizer |
| Naphtha Reformer |
| $H_2SO_4$ Plant |
| Compressor/Gasholder |

With the coal-fired boilers, both high and low sulfur coals were analyzed. However, only one set of control levels are considered (80 percent $NO_x$, 85 percent $SO_x$) and therefore, it is not possible to present costs as a function of control level as is done in the $NO_x$-only section. Instead, the costs are plotted against flue gas flow rate to show the effect of unit size on cost. The results for the residual oil-fired boiler are presented in tabular form, but not plotted since only one boiler and control level are considered.

### 4.2.2 Control Costs for Coal-Fired Boilers

The equipment listed in the previous table is sized based on material balances performed for each case. These balances are presented in Appendix 3. Detailed breakdowns of both capital and operating costs are presented in Appendices 6 and 7, respectively.

The annualized costs for the standard boilers considered are presented in Table 4.2.2-1 and plotted in Figure 4.2.2-1. The costs are significantly higher than those for the $NO_x$-only processes because the additional requirement of $SO_2$ removal necessitates the use of small processing units for $H_2$ production and $SO_2$ workup. In a real world situation where several industrial boilers operate at a single location, it will be possible to reduce costs by having large, central units for $H_2$ production and $SO_2$ workup. This option is not considered here since the cost impact is a function of the total number of boilers serviced by the central facilities and this is entirely site specific.

The cost to retrofit such a process can be calculated from the data presented in Section 4.1.2. Depending on the modifications required by the retrofit, the additional cost will be increased by an amount equivalent to 25 to 120 percent of the cost of an average $NO_x$-only system. The cost of special equipment necessary for $SO_2$ processing is not affected by a retrofit application.

### 4.2.3 Control Costs for the Oil-Fired Boiler

The equipment items necessary to treat flue gas from the residual oil-fired boiler are the same as for the coal-fired boilers. The annualized cost of the dry $NO_x/SO_x$ process applied to the residual oil-fired standard boiler

TABLE 4.2.2-1. COSTS OF $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Annual Costs | | Impacts* |
|---|---|---|---|---|---|---|---|
| Heat input | | | | | | | % increase in costs over uncontrolled boiler |
| MW | (MBtu/hr) | Type | Control efficiency (%) | Coal† | $/J/S | ($/MBtu/hr) | |
| 8.8 | (30) | Package Watertube Underfeed Stoker | 90% $NO_x$<br>85% $SO_x$ | HSE | 0.0811 | (23,690) | 75 |
| | | | | LSW | 0.0527 | (15,400) | 49 |
| 58.6 | (200) | Field Erected Watertube Pulverized Coal | 90% $NO_x$<br>85% $SO_x$ | HSE | 0.0309 | (9,025) | 42 |
| | | | | LSW | 0.0153 | (4,468) | 21 |

*Based only on Annual Costs

†HSE = High sulfur eastern coal (3.5% S)

LSW = Low sulfur western coal (0.6% S)

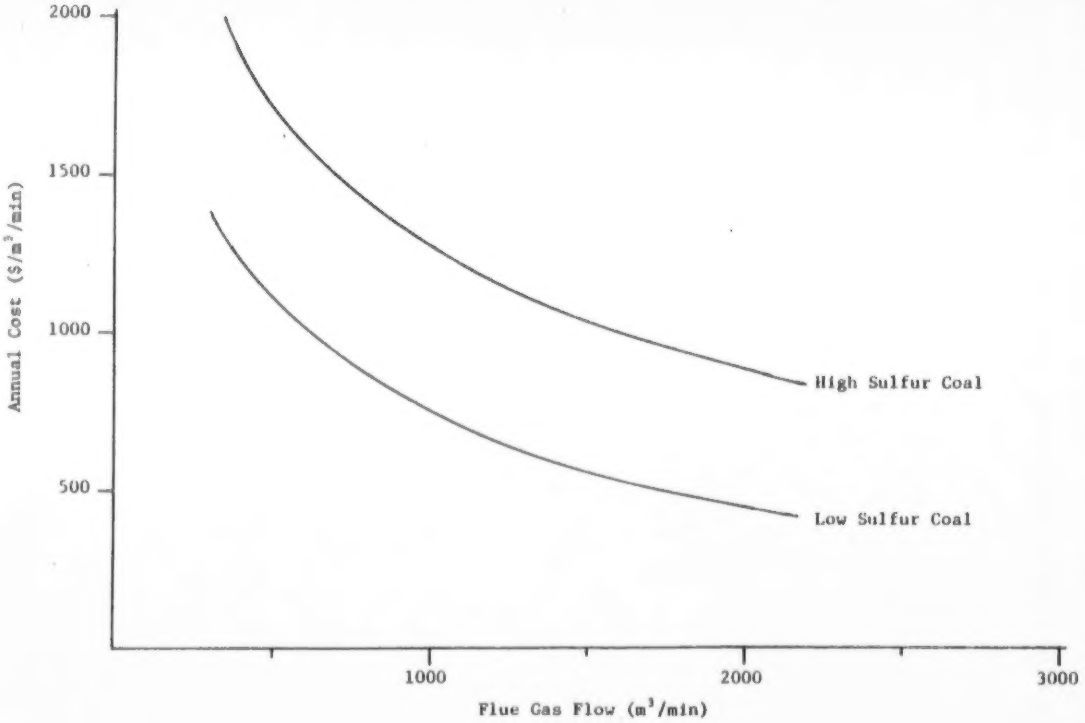

Figure 4.2.2-1. Annual cost of parallel flow SCR NO$_x$/SO$_x$ FGT for coal-fired boilers

is shown in Table 4.2.3-1. Detailed capital and annual costs are presented in detail for this case in the appendices. Retrofit considerations for this case are similar to those for the coal cases. The increased cost of a retrofit will be increased by the same dollar amount as the cost increase incurred when a similarly sized $NO_x$-only unit is retrofitted.

TABLE 4.2.3-1. COSTS OF THE DRY $NO_x/SO_x$ CONTROL TECHNIQUE FOR THE RESIDUAL OIL-FIRED BOILER

| Standard boiler |  |  |  | Annual costs |  | Impact* % increase in costs over uncontrolled boiler |
|---|---|---|---|---|---|---|
| Heat input |  |  |  |  |  |  |
| MW | (MBtu/hr) | Type | Control Level | $/J/S | ($/MBtu/hr) |  |
| 44 | (150) | Package Watertube | 90% $NO_x$<br>85% $SO_x$ | 0.0249 | (7280) | 43 |

*Based only on Annual Costs

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# SECTION 5

# ENERGY IMPACT

## 5.1 $NO_x$-ONLY SYSTEMS

### 5.1.1 Introduction

The three types of control systems selected in Section III for further comparison are analyzed with respect to energy requirements. All of the control systems are basically similar in principle and differ mainly in the design parameters. There are also a few differences in equipment requirements. Energy consumption steps considered in this analysis are listed in Table 5.1.1-1 for each of the control systems considered. It was assumed that flue gas could be taken from the boiler between the economizer and air heater at a temperature of 375°C. This eliminates the need for flue gas heating and heat exchange equipment. Since the hot flue gas is returned to the boiler upstream of the air heater, there is no loss in boiler efficiency.

Energy consumption was calculated using the design information and standard engineering principles. Design information from a variety of process developers was compared and used to generate a range of values or specific values. A range of values was determined for design parameters which changed with control level. Specific values for analysis were chosen from this range based on the level of control being considered, *e.g.* for 70% control a value at the lower end of the range was used. Design data used in this analysis is presented in Table 5.1.1-2.

TABLE 5.1.1-1. AREAS OF ENERGY CONSUMPTION IN $NO_x$ FGT SYSTEMS[1,2]

| $NO_x$ FGT system | Energy consumption step (equipment) | Type of energy consumed |
|---|---|---|
| Parallel Flow SCR | Reactor Draft Loss (Fan) | Electrical |
| | Liquid $NH_3$ Transfer (Pump) | Electrical |
| | $NH_3$ Vaporization (Vaporizer) | Steam |
| | $NH_3$ Dilution | Steam |
| Moving Bed SCR | Reactor Draft Loss (Fan) | Electrical |
| | Liquid $NH_3$ Transfer (Pump) | Electrical |
| | Catalyst Screening & Transfer (Elevator) | Electrical |
| | Baghouse Draft Loss (Blower) | Electrical |
| | $NH_3$ Vaporization (Vaporizer) | Steam |
| | $NH_3$ Dilution | Steam |
| Fixed Packed Bed SCR | Reactor Draft (Fan) | Electrical |
| | Liquid $NH_3$ Transfer (Pump) | Electrical |
| | $NH_3$ Vaporization (Vaporizer) | Steam |
| | $NH_3$ Dilution | Steam |
| | Soot Blowing-Distillate Oil Boiler Only | Steam |

TABLE 5.1.1-2. RANGE OF DESIGN PARAMETERS USED FOR ENERGY IMPACT CALCULATIONS[1,2,3]

| Parameter | Range or specific values used | | |
|---|---|---|---|
| | Parallel flow | Moving bed | Fixed packed bed |
| Space velocity | 3000-5000 | 6000-10000 | 6000-10000 |
| $NH_3$:$NO_x$ mole ratio | 0.7-1.0 | 0.7-1.0 | 0.7-1.0 |
| Dilution ratio (moles steam/mole $NH_3$) | 5:1 | 5:1 | 5:1 |
| Dilution steam pressure | 30 psig | 30 psig | 30 psig |
| Flue gas temperature | 375°C | 375°C | 375°C |
| Pressure drop | 80-160 $mmH_2O$ | 40-80 $mmH_2O$ | ~125 $mmH_2O$ |
| Catalyst type | Square honeycomb | Ring | Spherical pellet |
| Void fraction of packed catalyst particles | 0.67-0.7 | 0.52 | 0.26 |

Steam was chosen as the $NH_3$ dilution gas because of its ease of application and safety considerations. Air, at 20:1 air:$NH_3$ mole ratio, can also be used as an $NH_3$ diluent. Its use requires a compressor or blower and a motor which are high maintenance items. Also, at dilution ratios less than 20:1 there is an explosion hazard. The optimum choice would ordinarily be made by comparing the operating costs of steam use versus the capital charges of the air handling equipment plus the operating costs of electricity. This optimization is beyond the scope of this study and is site specific.

The analyses conducted in this study assumed that the boilers were operated constantly at full load and, therefore, had constant flue gas temperatures. However, it is possible that the boiler may experience large and frequent load swings which result in a variable flue gas temperature. FGT systems in this service will require flue gas heating in order to maintain sufficiently high temperatures. Temperature control can be accomplished by either a heater or a slipstream around the economizer. The heater will effectively decouple the FGT system from the boiler and does not require flow control of a flue gas slipstream. The economizer bypass will not derate the boiler since it will only be required during low load situations. Energy usage calculations were not made for either of these approaches since the amount of heating necessary is likely to be different for each boiler application.

### 5.1.2 Energy Impact of Controls for Coal-Fired Boilers

This subsection presents the results of calculations on the energy requirements of the candidate control systems applied to the standard boilers. One simplification was made in order to reduce the number of cases necessary for consideration and that is that only one coal was analyzed for each boiler. The justification for this simplification is presented below.

The result of the energy impact analyses indicate that the most significant energy consumption occurs in the fan required to overcome the reactor pressure drop and $NH_3$ dilution by steam. Coal sulfur content does not significantly affect the fan requirements which are a function of flue gas flow rate and control level. $NH_3$ dilution steam is affected; however, energy consumption of this step is approximately a third of the fan requirement. This is illustrated in an example calculation in Table 5.1.2-1. As can be seen the sulfur content of the coal does not significantly affect the total energy requirements especially when compared to the effect of control level.

The low sulfur coal was used for the analyses since the $NO_x$ emissions were somewhat higher and, therefore, energy usage for the other coals will not exceed those presented here.

Also, SIP control levels were not considered since in cases where control is required, it can be achieved through use of combustion modifications. The typical SIP control levels are shown in Table 5.1.2-2.

TABLE 5.1.2-2. SIP CONTROL LEVELS[4]

| Fuel | $NO_x$ emissions, $\frac{lb}{10^6\ Btu}$ | SIP level, $\frac{lb}{10^6\ Btu}$ | Required control efficiency |
|---|---|---|---|
| 3.5% S Coal | 0.64 | 0.7 | 0 |
| 0.9% S Coal | 0.55 | 0.7 | 0 |
| 0.6% S Coal | 0.78 | 0.7 | 10% |

Material balances were performed for each of the 7 cases considered for the coal-fired standard boilers. The results of these calculations appear in Appendices 3, 4, and 5. These results were used to calculate energy requirements of the control systems and an example calculation is presented in Appendix 8.

The results of the energy requirement calculations are presented in Tables 5.1.2-3 through 5.1.2-6. Each table represents one standard boiler and all control types and levels are included. It should be noted that the megawatt values shown for electrical usage are thermal megawatts and not

TABLE 5.1.2-1. RELATIVE SIGNIFICANCE OF PARAMETERS CONSIDERED IN ENERGY ANALYSIS
Example: Pulverized Coal Boiler, 90% Control, Parallel Flow SCR

| a) Effect of sulfur content | Energy usage (MW thermal) | | |
|---|---|---|---|
| | 0.6% S coal (187.56 lb $NO_x$/hr) | 0.9% S coal (130.50 lb $NO_x$/hr) | 3.5% S coal (152.46 lb $NO_x$/hr) |
| Energy Consumer | | | |
| Fan | 0.91 | 0.88 | 0.88 |
| Liquid NH Pump | 0.00373 | 0.00373 | 0.00373 |
| NH Vaporizer | 0.0383 | 0.0275 | 0.0325 |
| NH Dilution Steam | 0.325 | 0.234 | 0.275 |
| Total | 1.275 | 1.145 | 1.191 |

10% Difference

7% Difference

| b) Effect of removal level | 90% removal | 70% removal |
|---|---|---|
| Total Energy Consumed (MW thermal) | 1.28 | 0.821 |

36% Difference

TABLE 5.1.2-3. ENERGY CONSUMPTION FOR $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Energy consumption | | | |
|---|---|---|---|---|---|---|---|---|
| Standard boiler | | | | | | | | |
| Heat input | | Type | Type and level of control | Control efficiency % | Energy types | Energy consumed by control device | | % increase in energy use over uncontrolled boiler |
| MW | (MBtu/hr) | | | | | $MW_t$ | (MBtu/hr) | |
| | | | SCR Parallel Flow | | Electrical | 0.161 | (0.549) | |
| 58.6 | (200) | Field Erected, Watertube, Pulverized Coal | Moderate | 70 | Steam | 0.0797 | (0.272) | 0.41 |
| | | | SCR Parallel Flow | | Electrical | 0.268 | (0.912) | |
| | | | Stringent | 90 | Steam | 0.108 | (0.364) | 0.64 |

TABLE 5.1.2-4. ENERGY CONSUMPTION FOR $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Energy consumption | | | |
|---|---|---|---|---|---|---|---|---|
| Standard boiler | | | | | | | | |
| Heat input | | Type | Type and level of control | Control efficiency % | Energy types | Energy consumed by control device | | % increase in energy use over uncontrolled boiler |
| MW | (MBtu/hr) | | | | | $MW_t$ | (MBtu/hr) | |
| 44 | (150) | Field Erected, Watertube, Spreader Stoker | SCR Parallel Flow | | Electrical | 0.126 | (0.428) | |
| | | | Intermediate | 80 | Steam | 0.0568 | (0.194) | 0.41 |

## TABLE 5.1.2-5. ENERGY CONSUMPTION FOR NO$_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Energy Consumption | | | |
|---|---|---|---|---|---|---|---|---|
| Standard Boiler Heat Input | | Type | Type & Level of Control | Control Efficiency % | Energy Types | Energy Consumed by Control Device | | % Increase in Energy Use Over Uncontrolled Boiler |
| MW | (MBtu/hr) | | | | | MW$_t$ | (MBtu/hr) | |
| 22 | (75) | Package<br>Watertube<br>Chaingrate | SCR Parallel Flow | | Electrical | 0.0408 | (0.139) | |
| | | | Moderate | 70 | Steam | 0.0253 | (0.0862) | 0.30 |
| | | | SCR Parallel Flow | | Electrical | 0.0505 | (0.172) | |
| | | | Intermediate | 80 | Steam | 0.0289 | (0.0988) | 0.36 |
| | | | SCR Parallel Flow | | Electrical | 0.0669 | (0.228) | |
| | | | Stringent | 90 | Steam | 0.0337 | (0.115) | 0.46 |

## TABLE 5.1.2-6. ENERGY CONSUMPTION FOR NO$_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | Energy consumption | | | |
|---|---|---|---|---|---|---|---|---|
| Standard boiler Heat input | | Type | Type and level of control | Control efficiency % | Energy types | Energy consumed by control device | | % increase in energy use over uncontrolled boiler |
| MW | (MBtu/hr) | | | | | MW$_t$ | (MBtu/hr) | |
| 8.8 | (30) | Package<br>Watertube<br>Underfeed<br>Stoker | SCR Parallel Flow | | Electrical | 0.0185 | (0.0630) | |
| | | | Intermediate | 80 | Steam | 0.0113 | (0.0387) | 0.34 |

electrical megawatts. The data appearing in these tables is summarized in Table 5.1.2-7. This data is plotted in Figures 5.1.2-1 through 5.1.2-4. Each figure represents one standard boiler and shows the effect of removal level on energy usage for both types of FGT candidate systems: parallel flow SCR and moving bed SCR. Figure 5.1.2-5 presents energy usage for all boilers and FGT systems as a percent of the boiler heat input.

In general, energy usage seems to increase with control level in a nonlinear manner. This is also true with regard to energy usage as a percent of boiler input. Also, it appears that more energy is used per mole of $NO_x$ controlled with the larger boilers. This increase with boiler size is not a physical phenomena of SCR systems but rather an idiosyncrasy of the reactor design scheme. In keeping reactor geometry consistent from boiler to boiler, the linear gas velocity (m/s) decreased slightly with boiler size resulting in a corresponding slight decrease in pressure drop. However, the pressure drops of all the standard boilers are in the range of commercial operations and the differences in energy usage as a percentage of boiler heat input of the standard boilers are not large.

The effect of this energy usage on economics is examined in Section IV.

Very little work has been done with SCR systems to affect reductions in energy consumption. Problem solving efforts thus far have been directed toward improving the reliability of operating units and applying the control techniques to coal-fired flue gas. It is likely that only after the processes have been applied and demonstrated on coal-fired units will the overall energy consumption be examined in detail. It should be noted that the SCR processes are the least energy intensive of all of the FGT control systems mentioned in Section II.[1,5]

There are two areas in which there is a potential for energy savings. These are control of excess air and $NH_3$ dilution. By using only as much excess air as necessary, the energy required for pressure drop will be reduced. This has a twofold effect. Not only is the flow through the

TABLE 5.1.2-7. SUMMARY OF ENERGY REQUIREMENTS FOR COAL-FIRED INDUSTRIAL BOILERS

| | Pulverized coal | | Spreader stoker | | Chaingrate | | Underfeed stoker | |
|---|---|---|---|---|---|---|---|---|
| | Total thermal kW | % of boiler heat input | Total thermal kW | % of boiler heat input | Total thermal kW | % of boiler heat input | Total thermal kW | % of boiler heat input |
| Parallel Flow SCR | | | | | | | | |
| 90% Removal | 376 | 0.64 | - | - | 101 | 0.46 | - | - |
| 80% Removal | - | - | 183 | 0.41 | 80 | 0.36 | 30 | 0.34 |
| 70% Removal | 241 | 0.41 | - | - | 66 | 0.30 | - | - |

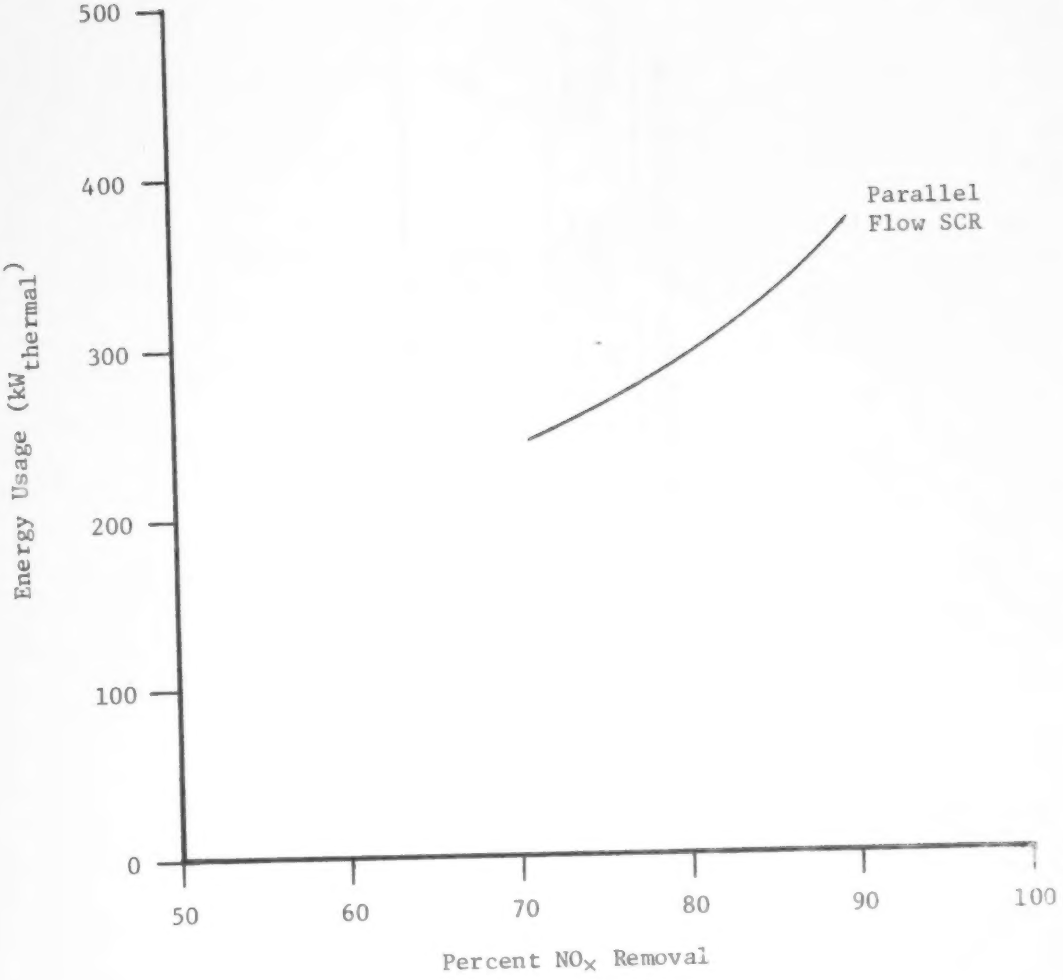


Figure 5.1.2-1. Energy usage of $NO_x$ control systems for pulverized coal standard boiler.

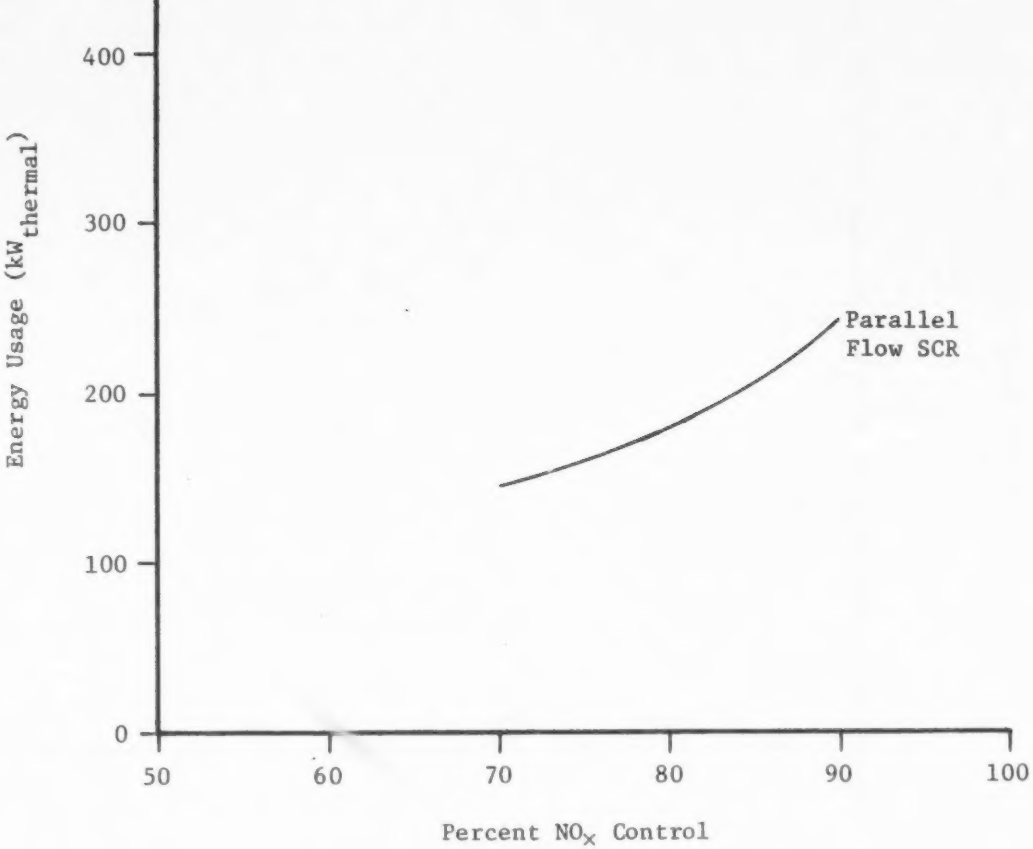


Figure 5.1.2-2. Energy usage of $NO_x$ control systems for spreader stoker standard boiler.

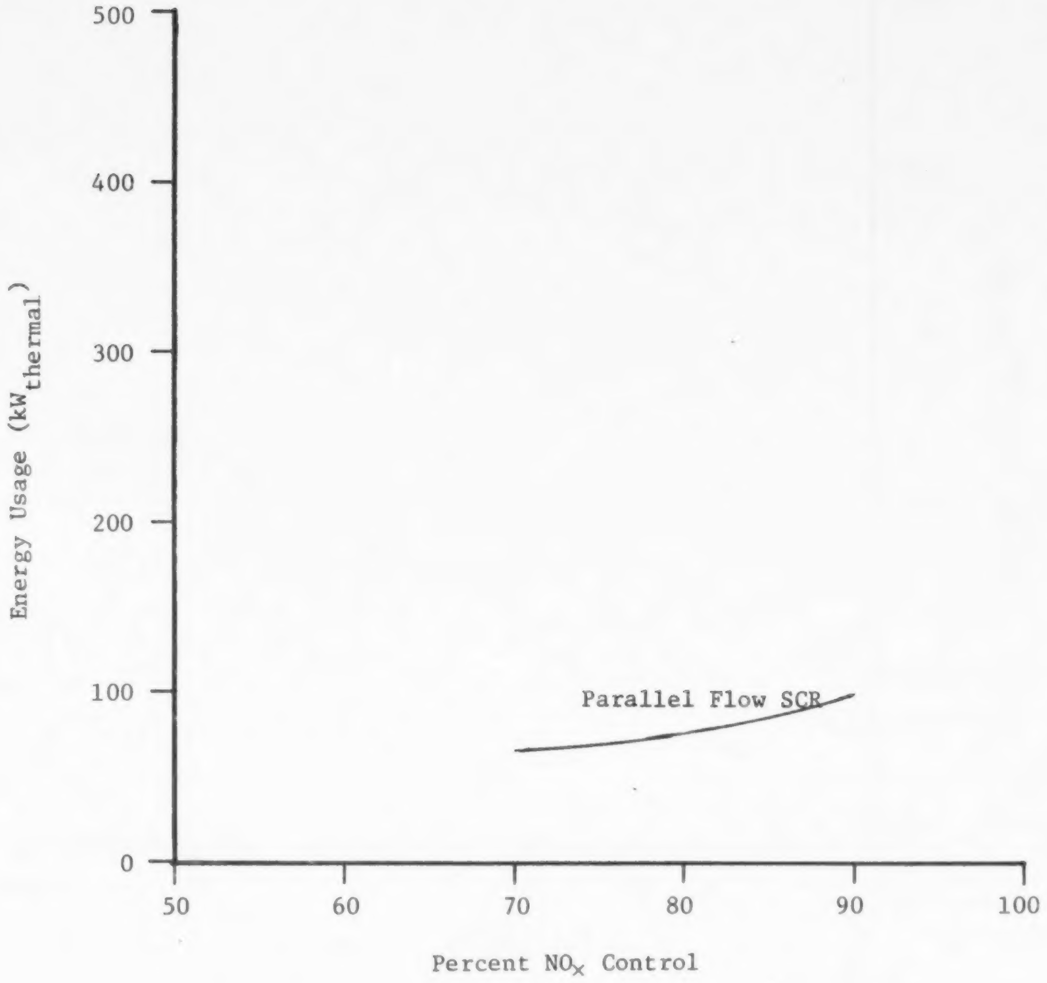


Figure 5.1.2-3. Energy usage of NO$_x$ control systems for chaingrate standard boiler.

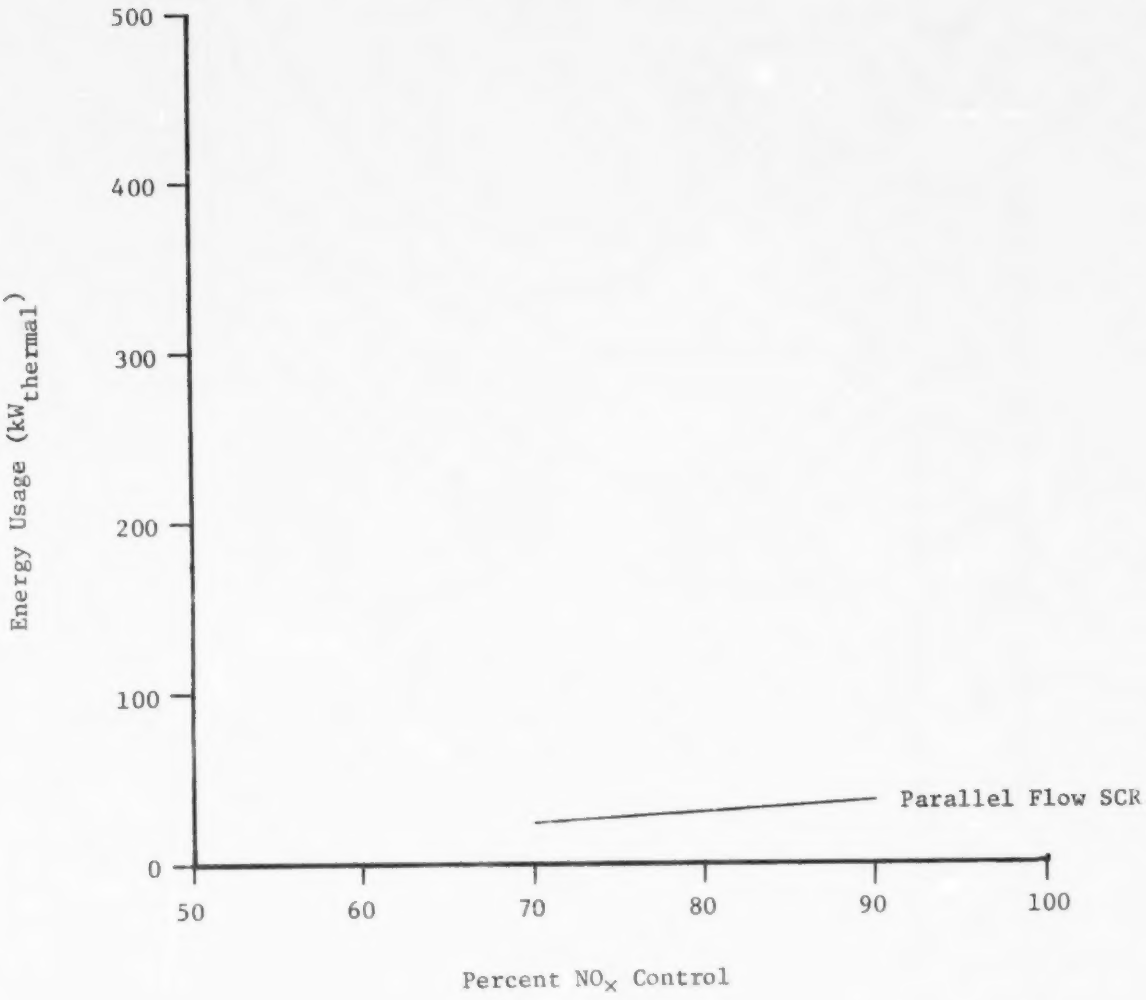


Figure 5.1.2-4. Energy usage of $NO_x$ control systems for underfeed stoker standard boiler.

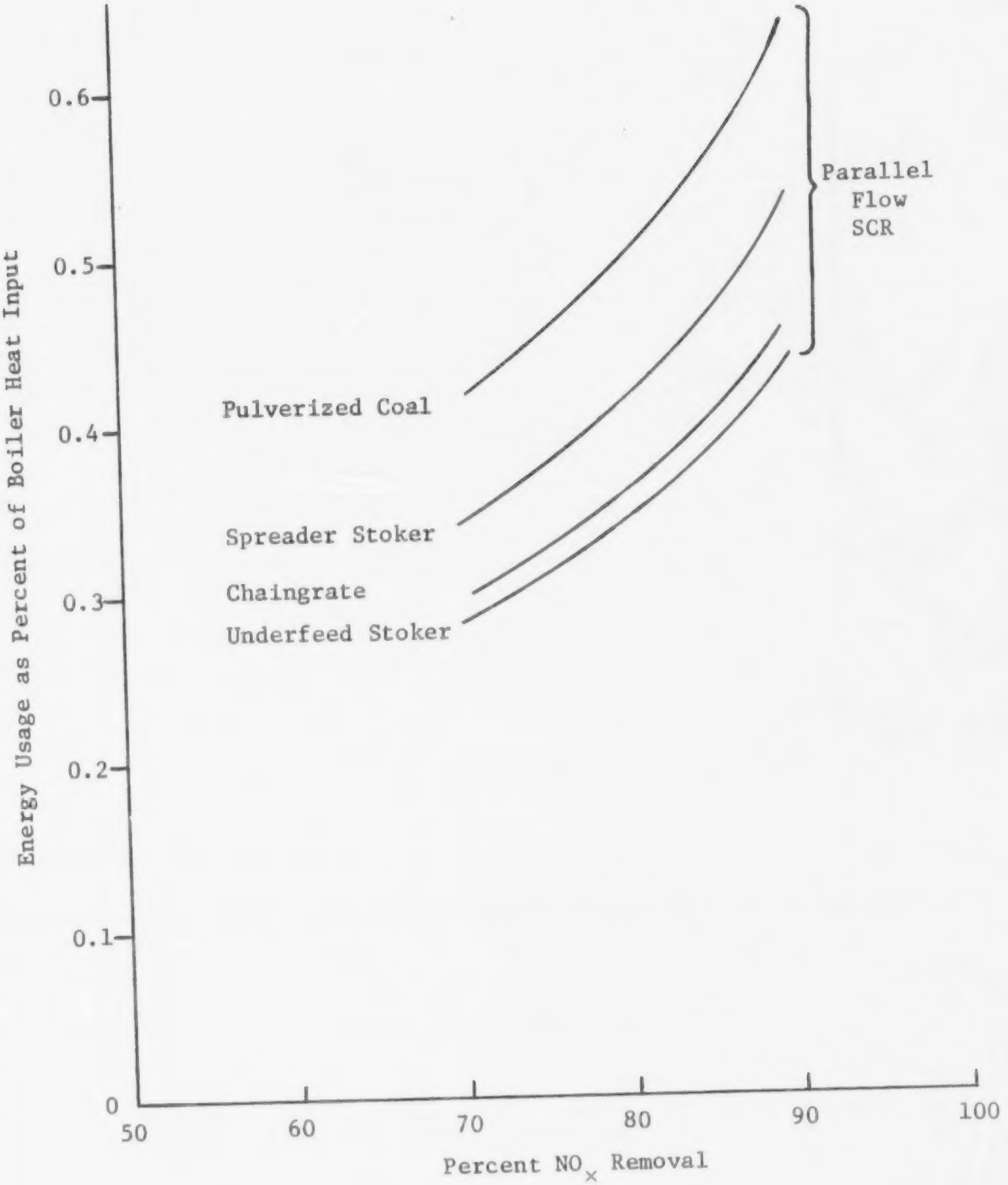


Figure 5.1.2-5. Energy usage of $NO_x$ control systems as percent of boiler heat input.

reactor reduced, but the required reactor volume itself is reduced by lower flue gas flow rates. It is likely that a boiler equipped with combustion modifications will utilize low excess air for $NO_x$ control. Energy consumption by $NH_3$ dilution might be reduced by using air instead of steam at a specific site. Use of air is less safe since some air:$NH_3$ mixtures can be explosive.

The energy impact of FGT controls applied to modified or reconstructed facilities (retrofit application) will be the same or greater than that for new facilities. If flue gas can be taken from the economizer of the existing boiler at 350-400°C and returned upstream of any existing heat exchange equipment, then there will be no additional energy impact.

If the flue gas is only available at a lower temperature (<350-400°C) then a heater will be required. The energy impact of the heater will depend on the temperature of the flue gas. If the temperature is that of the outlet gas of the standard boilers (approximately 180°C), calculations indicate that energy requirement would be more than tripled even if heat exchange equipment is used to recover 85% of the energy supplied by the heater. The heater will probably be oil-fired for ease of control.

These results indicate that, on retrofit applications, there is a considerable energy incentive to obtain the flue gas at the necessary reaction temperature in order to avoid flue gas heating. Other energy impacts would be the same as those for new facilities.

### 5.1.3 Energy Impact of Controls for Oil-Fired Boilers

In this subsection, the results of energy impact calculations for the candidate FGT systems as applied to the standard oil-fired boilers are presented. The combinations considered are

| Boiler Size, $MW_t$ | Fuel | FGT System |
|---|---|---|
| 8.8, 44 | Residual Oil | Parallel Flow SCR |
| 8.8, 44 | Residual Oil | Moving Bed SCR |
| 4.4, 44 | Distillate Oil | Fixed Packed Bed SCR |

Also, two levels of control are considered for each combination.

The first step in performing this energy impact analysis was to calculate general material balances. The result of these balances were used to determine energy requirements for each process step. Energy consuming steps and the types of energy used were presented earlier in Table 5.1.1-1. All calculations are similar to the example case presented in Appendix 8.

The results of these calculations are presented in Tables 5.1.3-1 and 5.1.3-2. The data in Table 5.1.3-1 represents energy consumption for residual oil-fired boilers. Two candidate systems and two levels of control are considered. Table 5.1.3-2 shows energy consumption for application of a fixed packed bed SCR process to the standard boiler firing distillate oil.

All energy values presented are on a thermal basis. Actual electrical usages have been converted to a heat input basis by assuming 10,000 Btu/hr per kW.

The data appearing in Tables 5.1.3-1 and 5.1.3-2 is summarized in Table 5.1.3-3 and is plotted in Figures 5.1.3-1 through 5.1.3-4. The first two figures show thermal energy usage as a function of $NO_x$ control for all fuels and control systems. The next two figures illustrate energy usage as a percent of boiler heat input for all cases.

Energy usage increases in a nonlinear manner with control level. The energy usage as a percent of boiler input is also nonlinear. On this basis the fixed packed bed SCR appears to be the most energy intensive and the moving bed SCR the least. It is difficult to draw any definite conclusions

## TABLE 5.1.3-1. ENERGY CONSUMPTION FOR $NO_x$ FGT CONTROL TECHNIQUES FOR RESIDUAL OIL-FIRED BOILERS

| System | | | | | Energy consumption | | | |
|---|---|---|---|---|---|---|---|---|
| Standard boiler | | | Type and level of control | Control efficiency % | Energy types | Energy consumed by control device | | % increase in energy use over uncontrolled boiler |
| Heat input | | Type | | | | | | |
| $MW_t$ | (MBtu/hr) | | | | | $MW_t$ | (MBtu/hr) | |
| 8.8 | (30) | Package Watertube | SCR Parallel Flow Moderate | 70 | Electrical<br>Steam | 0.011<br>0.0071 | (0.0367)<br>(0.0242) | 0.20 |
| | | | SCR Parallel Flow Stringent | 90 | Electrical<br>Steam | 0.018<br>0.0095 | (0.0597)<br>(0.0323) | 0.31 |
| | | | SCR Moving Bed Moderate | 70 | Electrical<br>Steam | 0.0094<br>0.0071 | (0.0322)<br>(0.0242) | 0.19 |
| | | | SCR Moving Bed Stringent | 90 | Electrical<br>Steam | 0.014<br>0.0095 | (0.0462)<br>(0.0323) | 0.26 |
| 44 | (150) | Package Watertube | SCR Parallel Flow Moderate | 70 | Electrical<br>Steam | 0.0813<br>0.0253 | (0.277)<br>(0.0864) | 0.24 |
| | | | SCR Parallel Flow Stringent | 90 | Electrical<br>Steam | 0.134<br>0.0337 | (0.458)<br>(0.115) | 0.38 |
| | | | SCR Moving Bed Moderate | 70 | Electrical<br>Steam | 0.0570<br>0.0253 | (0.195)<br>(0.0864) | 0.19 |
| | | | SCR Moving Bed Stringent | 90 | Electrical<br>Steam | 0.0918<br>0.0337 | (0.314)<br>(0.115) | 0.29 |

TABLE 5.1.3-2. ENERGY CONSUMPTION FOR $NO_x$ FGT CONTROL TECHNIQUES FOR DISTILLATE OIL-FIRED BOILERS

| System | | | | | Energy consumption | | | |
|---|---|---|---|---|---|---|---|---|
| Standard boiler | | | | | | | | |
| Heat input | | Type | Type and level of control | Control efficiency % | Energy types | Energy consumed by control device | | % increase in energy use over uncontrolled boiler |
| $MW_t$ | (MBtu/yr) | | | | | $MW_t$ | (MBtu/yr) | |
| 4.4 | (15) | Package Firetube Scotch | SCR Fixed Packed Bed Moderate | 70 | Electrical<br>Steam | 0.00994<br>0.00697 | (0.0339)<br>(0.0238) | 0.38 |
| | | | SCR Fixed Packed Bed Stringent | 90 | Electrical<br>Steam | 0.0158<br>0.00888 | (0.0539)<br>(0.0302) | 0.56 |
| 44 | (150) | Package Watertube | SCR Fixed Packed Bed Moderate | 70 | Electrical<br>Steam | 0.200<br>0.0734 | (0.682)<br>(0.251) | 0.62 |
| | | | SCR Fixed Packed Bed Stringent | 90 | Electrical<br>Steam | 0.121<br>0.0706 | (0.414)<br>(0.241) | 0.44 |

## TABLE 5.1.3-3. SUMMARY OF ENERGY REQUIREMENTS FOR OIL-FIRED INDUSTRIAL BOILERS

| | 8.8 $MW_t$ Residual oil | | 44 $MW_t$ Residual oil | | 4.4 $MW_t$ Distillate oil | | 44 $MW_t$ Distillate oil | |
|---|---|---|---|---|---|---|---|---|
| | Total thermal kW | % of boiler heat input | Total thermal kW | % of boiler heat input | Total thermal kW | % of boiler heat input | Total thermal kW | % of boiler heat input |
| Parallel Flow SCR | | | | | | | | |
| 90% Removal | 27 | 0.31 | 168 | 0.38 | * | | * | |
| 70% Removal | 18 | 0.20 | 107 | 0.24 | * | | * | |
| Moving Bed SCR | | | | | | | | |
| 90% Removal | 23 | 0.26 | 126 | 0.29 | * | | * | |
| 70% Removal | 17 | 0.19 | 82 | 0.19 | * | | * | |
| Fixed Packed Bed SCR | | | | | | | | |
| 90% Removal | * | | * | | 25 | 0.56 | 274 | 0.62 |
| 70% Removal | * | | * | | 17 | 0.38 | 192 | 0.44 |

*Not considered as a candidate system.

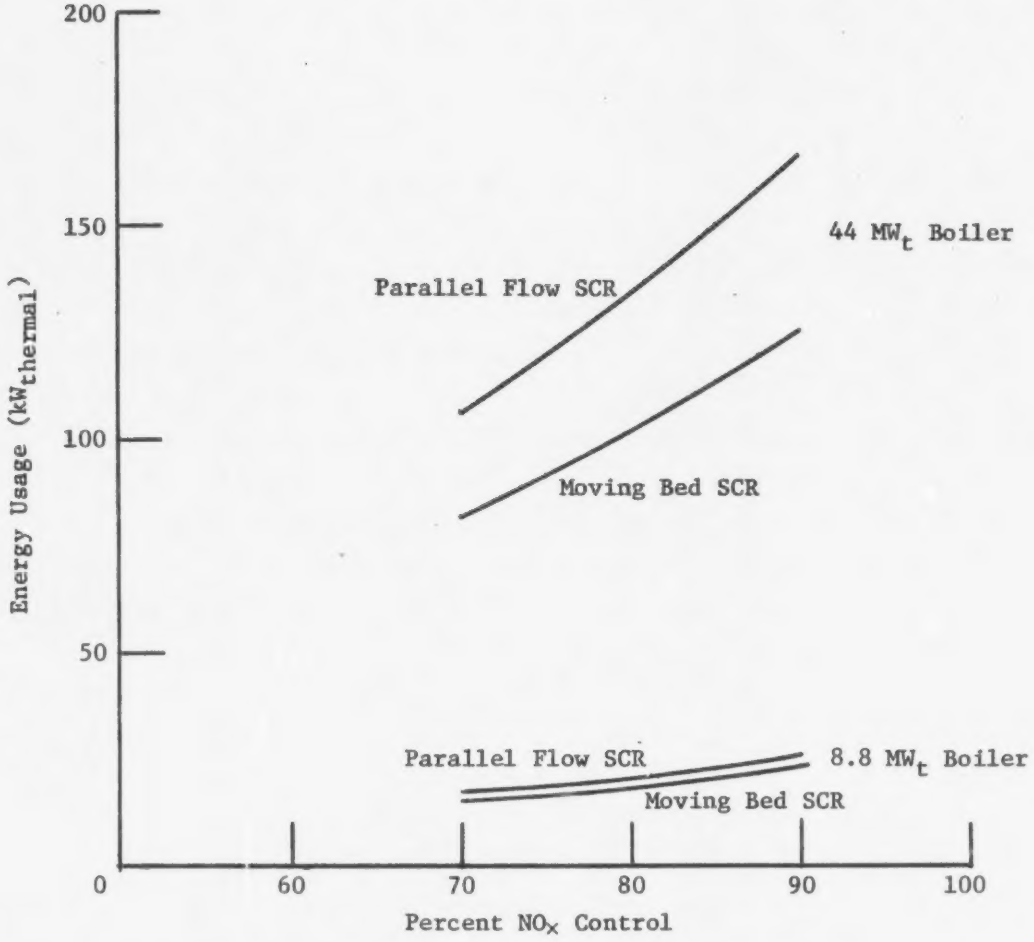


Figure 5.1.3-1. Energy usage of NO$_x$ control systems for residual oil-fired standard boilers.

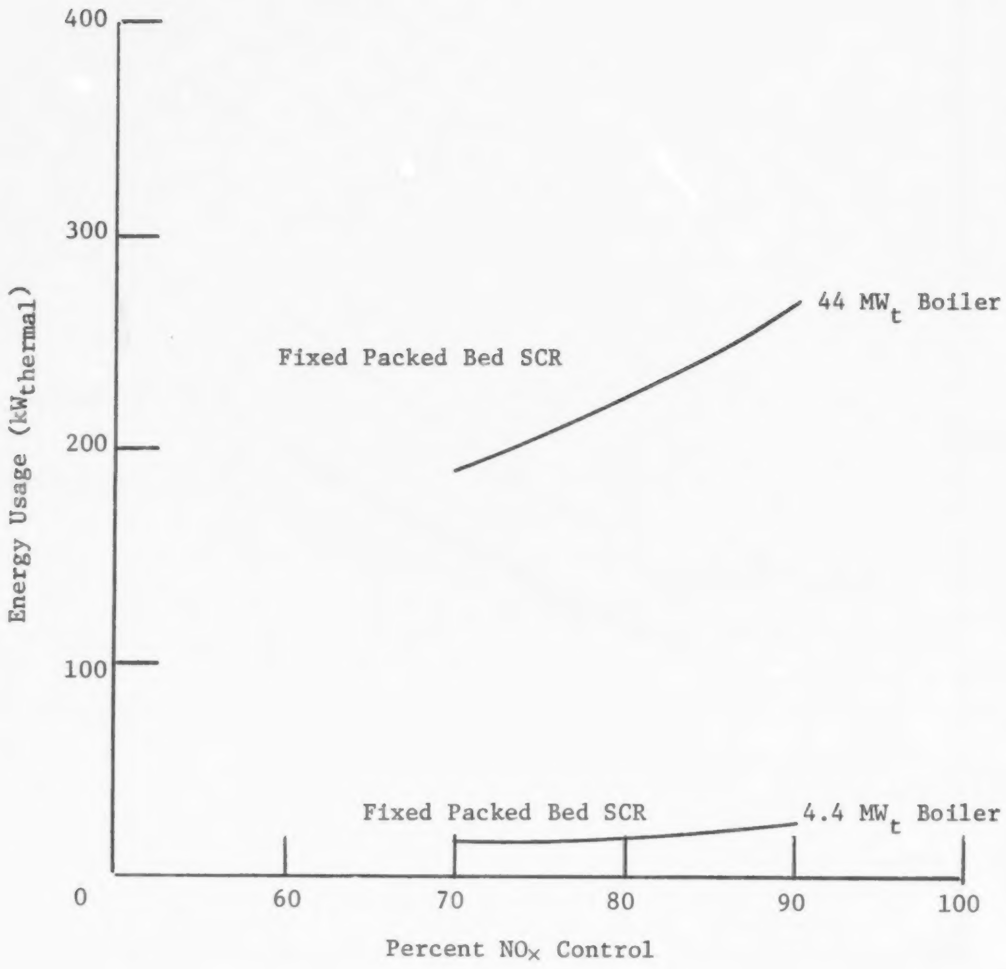


Figure 5.1.3-2. Energy usage of $NO_x$ control systems for distillate oil-fired boilers.

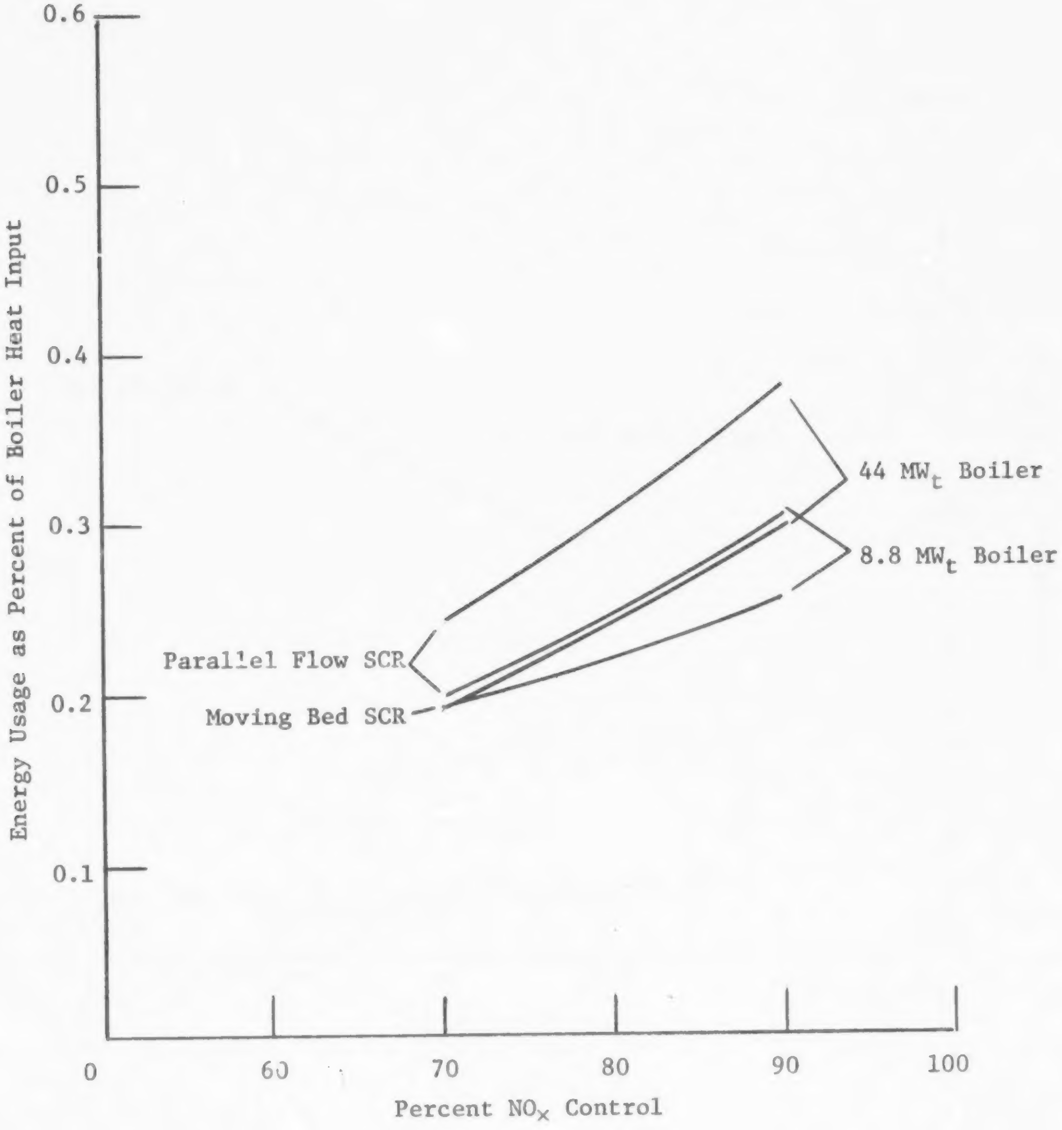


Figure 5.1.3-3. Energy usage of $NO_x$ control systems applied to residual oil-fired boilers as percent of boiler heat input.

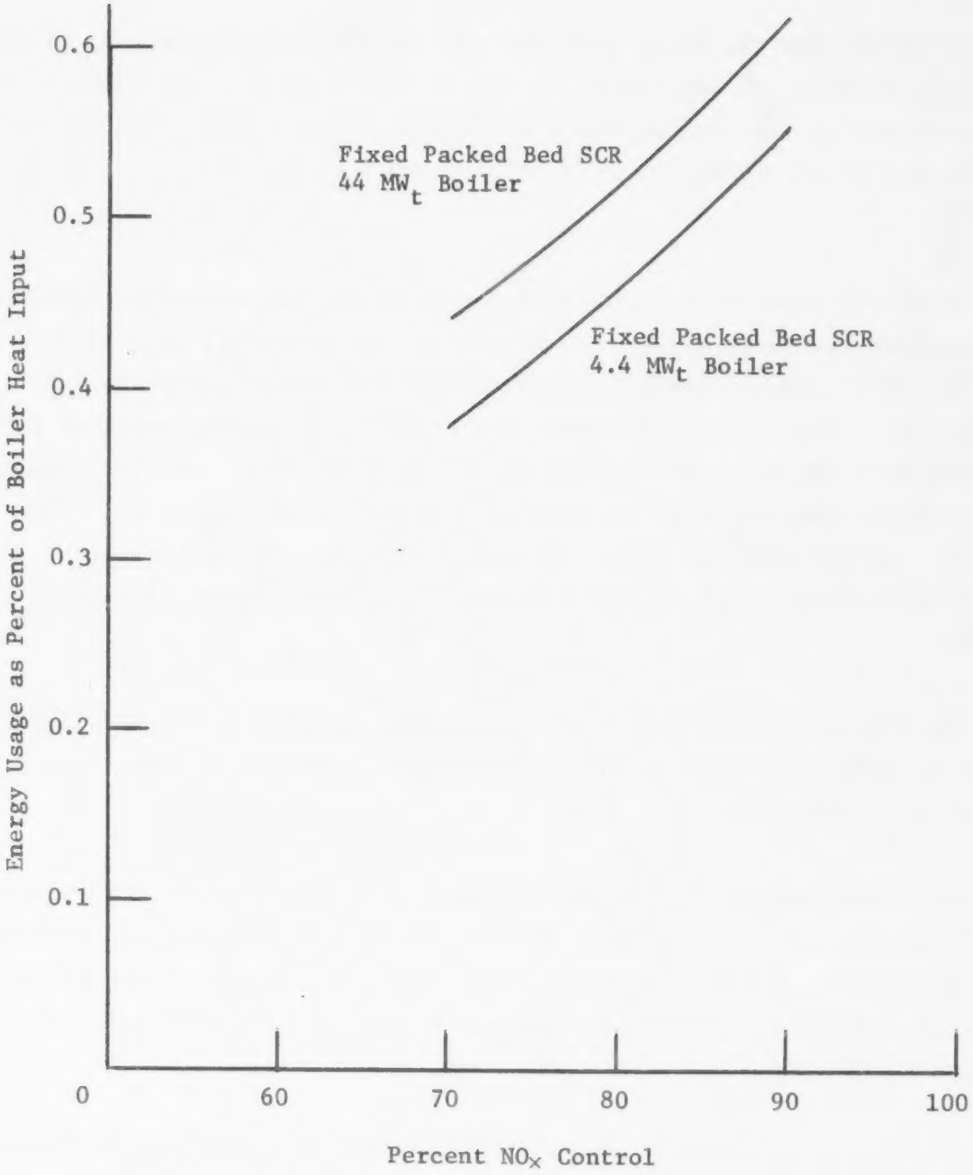


Figure 5.1.3-4. Energy usage of $NO_x$ control systems applied to distillate oil-fired boilers as percent of boiler heat input.

when comparing the two fuels since the size of the standard boilers is different by an order of magnitude. It can be said then, for the residual oil case, the moving bed systems are less energy intensive than parallel flow systems due to the moving beds[1] lower pressure drop across the length of the reactor.

It is not clear as to whether or not these systems have been optimized with respect to energy usage. The technology is relatively new and problem solving efforts are probably directed toward improving reliability and operability. It does seem possible that there is an optimum catalyst size and reactor volume that would minimize the pressure drop. Another potential method of lowering the pressure drop is to minimize the excess air. This reduces both the required reactor volume and the $\Delta P$. It is likely that a boiler equipped with low $NO_x$ burners will utilize low excess air for $NO_x$ control.

$NH_3$ dilution by air instead of steam might possibly use less energy. There is, however, a safety aspect to consider since some air/$NH_3$ mixtures (<20:1) are explosive.

The energy impact of FGT controls applied to modified or reconstructed facilities (retrofit application) will be the same or greater than that for new facilities. If flue gas can be taken from the economizer of the existing boiler at 350-400°C and returned upstream of any existing heat exchange equipment, then there will be no additional energy impact.

If the flue gas is only available at a lower temperature (<350-400°C) then a heater will be required. The energy impact of the heater will depend on the temperature of the flue gas. If the temperature is that of the outlet gas of the standard boilers (approximately 180°C) calculations indicate that energy requirement would be more than tripled even if heat exchange equipment is used to recover 85% of the energy supplied by the heater. The heater will probably be oil-fired for ease of control.

These results indicate that, on retrofit applications, there is a considerable energy incentive to obtain the flue gas at the necessary reaction temperature in order to avoid flue gas heating. Other energy impacts would be the same as those for new facilities.

### 5.1.4 Energy Impact of Controls for Natural Gas-Fired Boilers

This subsection presents the results of energy and material balance for natural gas-fired industrial boilers. For new facilities two standard boilers and one $NO_x$ FGT system is considered. Results are presented for two levels of control.

The data presented is the result of several calculations. First, material balances were performed and the necessary equipment sized. Then, knowing the equipment size and flow rates, it was possible to calculate energy usage for each process step.

The results of these calculations are presented in Table 5.1.4-1. Both thermal energy requirements and energy requirements as a percentage of boiler heat input are shown. The candidate system for natural gas-fired boilers is fixed packed bed SCR. For the calculations, it is assumed that flue gas is available from the boiler economizer at 375°C and can be returned upstream of the air heater. Therefore, no energy is necessary for flue gas heating.

The data appearing in Table 5.1.4-1 is summarized in Table 5.1.4-2 and plotted in Figures 5.1.4-1 and 5.1.4-2. Figure 5.1.4-1 presents total energy usage and Figure 5.1.4-2 shows the energy usage as a percent of boiler heat input.

There are some areas of potential energy usage reduction. The catalyst particle size and reactor volume may be optimized to minimize reactor pressure drop. Reduction of excess air may also reduce the pressure drop and this may be standard practice on boilers with low $NO_x$ burners. It may be more

## TABLE 5.1.4-1. ENERGY CONSUMPTION FOR $NO_x$ FGT CONTROL TECHNIQUES FOR NATURAL GAS-FIRED BOILERS

| System | | | | | | | | |
|---|---|---|---|---|---|---|---|---|
| Standard boiler | | | | | | | | |
| Heat input | | | | | | | | |
| $MW_t$ | (MBtu/hr) | | Type and level of control | Control efficiency % | Energy types | Energy consumed by control device $MW_t$ | Energy consumed by control device (MBtu/yr) | % increase in energy use over uncontrolled boiler |
| 4.4 | (15) | Package<br>Firetube<br>Scotch | SCR Fixed Packed Bed<br>Moderate | 70 | Electrical<br>Steam | 0.0108<br>0.00106 | (0.0369)<br>(0.00363) | 0.27 |
| | | | SCR Fixed Packed Bed<br>Stringent | 90 | Electrical<br>Steam | 0.0173<br>0.00133 | (0.0590)<br>(0.00455) | 0.42 |
| 44 | (150) | Package<br>Watertube | SCR Fixed Packed Bed<br>Moderate | 70 | Electrical<br>Steam | 0.123<br>0.0110 | (0.421)<br>(0.0345) | 0.30 |
| | | | SCR Fixed Packed Bed<br>Stringent | 90 | Electrical<br>Steam | 0.203<br>0.0142 | (0.692)<br>(0.0483) | 0.49 |

## TABLE 5.1.4-2. SUMMARY OF ENERGY REQUIREMENTS FOR NATURAL GAS-FIRED BOILERS

| | Natural gas | |
|---|---|---|
| | Total thermal kW | % of boiler heat input |
| 4.4 $MW_t$ Boiler<br>Fixed Packed Bed SCR | | |
| 90% Removal | 19 | 0.42 |
| 70% Removal | 12 | 0.27 |
| 44 $MW_t$ Boiler<br>Fixed Packed Bed SCR | | |
| 90% Removal | 217 | 0.49 |
| 70% Removal | 134 | 0.30 |

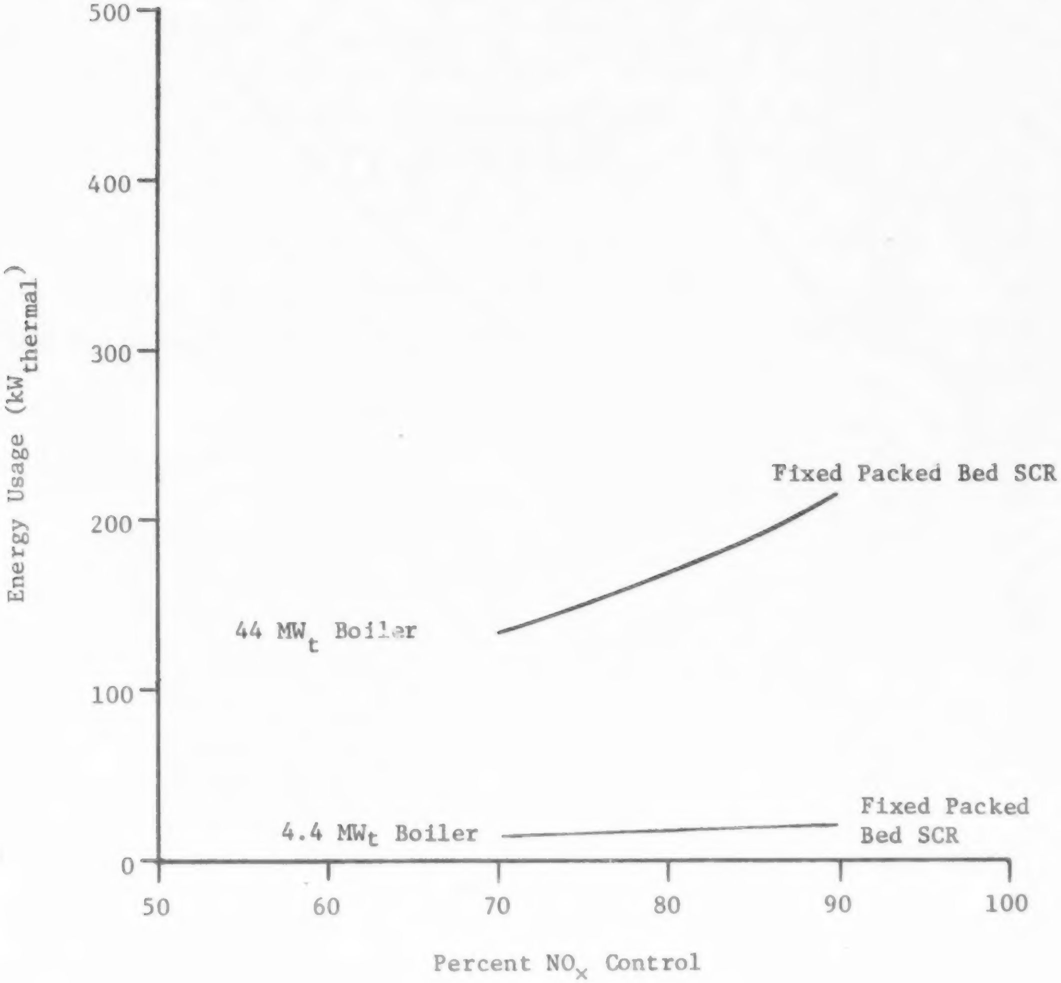
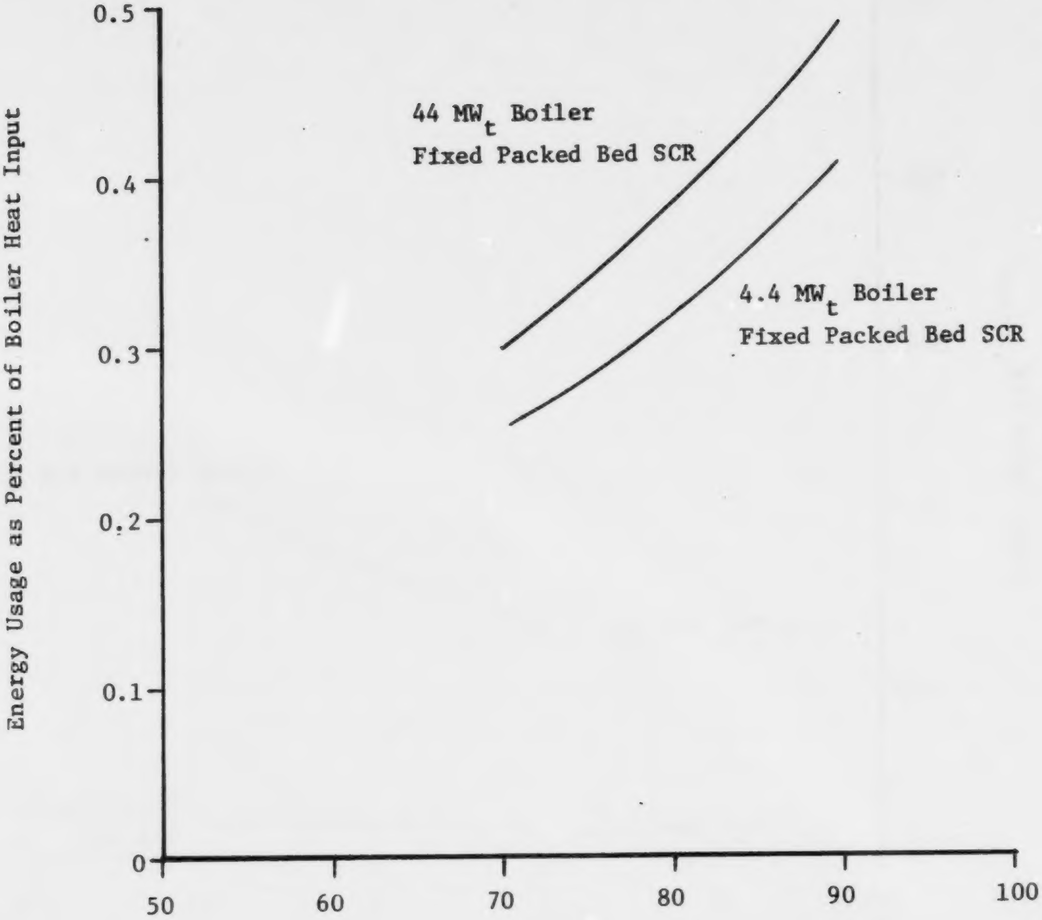


Figure 5.1.4-1. Energy usage of $NO_x$ control systems for natural gas-fired standard boiler.



Percent $NO_x$ Control

Figure 5.1.4-2. Energy usage of $NO_x$ control systems as percent of boiler heat input.

energy efficient to use air instead of steam for $NH_3$ dilution, however, there is an explosion hazard with some air:$NH_3$ mixtures (<20:1).

The energy impact of FGT controls applied to modified or reconstructed facilities (retrofit application) will be the same or greater than that for new facilities. If flue gas can be taken from the economizer of the existing boiler at 350-400°C and returned upstream of any existing heat exchange equipment, then there will be no additional energy impact.

If the flue gas is only available at a lower temperature (<350-400°C) then a heater will be required. The energy impact of the heater will depend on the temperature of the flue gas. If the temperature is that of the outlet gas of the standard boilers (approximately 180°C) calculations indicate that energy requirement would be more than tripled even if heat exchange equipment is used to recover 85% of the energy supplied by the heater. The heater will probably be oil-fired for ease of control.

These results indicate that, on retrofit applications, there is a considerable energy incentive to obtain the flue gas at the necessary reaction temperature in order to avoid flue gas heating. Other energy impacts would be the same as those for new facilities.

## 5.2 $NO_x/SO_x$ SYSTEMS

### 5.2.1 Introduction

This section considers the energy impacts associated with applying the UOP $NO_x/SO_x$ FGT system to three industrial boilers. The combinations analyzed are presented in Table 5.2.1-1.

The $NO_x/SO_x$ system has several more energy inputs than the $NO_x$ only systems; however, much of this energy is recovered by the air preheater resulting in an energy credit. The areas of energy utilization are shown in Table 5.2.1-2.

TABLE 5.2.1-1. $NO_x/SO_x$ FGT/BOILER COMBINATIONS ANALYZED FOR ENERGY IMPACT

| $NO_x/SO_x$ System | Boiler | Fuel* | Control Level $NO_x$ | Control Level $SO_x$ |
|---|---|---|---|---|
| UOP | Pulverized Coal | LSW<br>HSE | 80 | 85 |
| UOP | Underfeed Stoker | LSW<br>HSE | 80 | 85 |
| UOP | Oil-Fired | Residual Oil | 80 | 85 |

* LSW = Low sulfur western coal (0.6% S)
HSE = High sulfur eastern coal (3.5% S)

TABLE 5.2.1-2. AREAS OF ENERGY UTILIZATION IN THE $NO_x/SO_x$ FGT SYSTEM

| Process Step | Type of Energy Consumed |
|---|---|
| Reactor Draft Loss (Fan) | Electrical |
| Liquid $NH_3$ Transfer (Pump) | Electrical |
| $NH_3$ Vaporization (Vaporizer) | Steam |
| $NH_3$ Dilution | Steam |
| Naphtha Reformer | Electrical, Steam, Fuel |
| Compressor/Gasholder | Electrical |
| $H_2SO_4$ Plant | Electrical, Steam |

For each case, a heat and material balance is performed and these are contained in Appendices 4 and 5. These are used to size the equipment and determine the energy requirements. These requirements are listed in tabular form in each section and summarized. Since only one removal level is considered, the energy usage is not plotted against removal level as in the $NO_x$-only section.

### 5.2.2 Energy Impact of $NO_x/SO_x$ Controls for Coal-Fired Boilers

Energy usage by these $NO_x/SO_x$ applications is fairly evenly divided among three energy types: electrical, steam and fuel. These data are presented in Tables 5.2.2-1 and 5.2.2-2. Also shown in the tables are the heat credits for energy recovered by the air preheater.

The net energy usage by the $NO_x/SO_x$ system is higher than that of the $NO_x$-only systems. When put on the basis of percent increase in energy over that of the uncontrolled boiler, the energy usage appears to be a function of the coal sulfur content, but not unit size. Removal level will also impact the energy usage; however, the magnitude of this impact is not known. Energy usage is summarized in Table 5.2.2-3 and plotted in Figure 5.2.2-1.

TABLE 5.2.2-3. SUMMARY OF ENERGY USAGE OF $NO_x/SO_x$ SYSTEMS APPLIED TO COAL-FIRED BOILERS

| Fuel | Pulverized coal | | Underfeed stoker | |
|---|---|---|---|---|
| | Thermal kW | % of boiler heat input | Thermal kW | % of boiler heat input |
| Low Sulfur Western Coal | 1,240 | 2.1 | 200 | 2.3 |
| High Sulfur Eastern Coal | 11,200 | 7.7 | 680 | 7.7 |

TABLE 5.2.2-1. ENERGY CONSUMPTION FOR $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL FIRED BOILERS

| System | | | | | | Energy Consumption | | | |
|---|---|---|---|---|---|---|---|---|---|
| Standard Boiler | | | Coal Type | Type and Level of Control | Control Efficiency (% $NO_x/SO_x$) | Energy Types | Energy Consumed By Control Device | | % Increase in Energy Use Over Uncontrolled Boiler |
| Heat Input | | Type | | | | | | | |
| $MW_t$ | (MBtu/hr) | | | | | | $MW_t$ | (MBtu/hr) | |
| 58.6 | 200 | Field Erected Watertube Pulverized Coal | High Sulfur Eastern Coal | SCR Parallel Flow Intermediate | 80/85 | Electrical | 9.45 | (32.25) | 7.7 |
| | | | | | | Steam | 2.79 | ( 9.52) | |
| | | | | | | Fuel | 5.18 | (17.69) | |
| | | | | | | Heat Credit | -(6.24) | -(21.3 ) | |
| 58.6 | 200 | | Low Sulfur Western Coal | SCR Parallel Flow Intermediate | 80/85 | Electrical | 0.941 | ( 3.21) | 2.1 |
| | | | | | | Steam | 0.703 | ( 2.40) | |
| | | | | | | Fuel | 1.09 | ( 3.72) | |
| | | | | | | Heat Credit | -(1.49) | -( 5.1) | |

TABLE 5.2.2-2. ENERGY CONSUMPTION FOR $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

| System | | | | | | Energy Consumption | | | |
|---|---|---|---|---|---|---|---|---|---|
| Standard Boiler | | | Coal Type | Type and Level of Control | Control Efficiency (% $NO_x/SO_x$) | Energy Types | Energy Consumed By Control Device | | % Increase in Energy Use Over Uncontrolled Boiler |
| Heat Input | | Type | | | | | | | |
| $MW_t$ | (MBtu/hr) | | | | | | $MW_t$ | (MBtu/hr) | |
| 8.8 | 30 | Package Watertube Underfeed | High Sulfur Eastern Coal | SCR Parallel Flow Intermediate | 80/85 | Electrical | 0.428 | (1.46 ) | 7.7 |
| | | | | | | Steam | 0.416 | (1.42 ) | |
| | | | | | | Fuel | 0.774 | (2.64 ) | |
| | | | | | | Heat Credit | -(0.935) | -(3.19 ) | |
| 8.8 | 30 | | Low Sulfur Western Coal | SCR Parallel Flow Intermediate | 80/85 | Electrical | 0.151 | (0.52) | 2.3 |
| | | | | | | Steam | 0.102 | (0.348) | |
| | | | | | | Fuel | 0.163 | (0.558) | |
| - | | | | | | Heat Credit | -(0.217) | -(0.74 ) | |

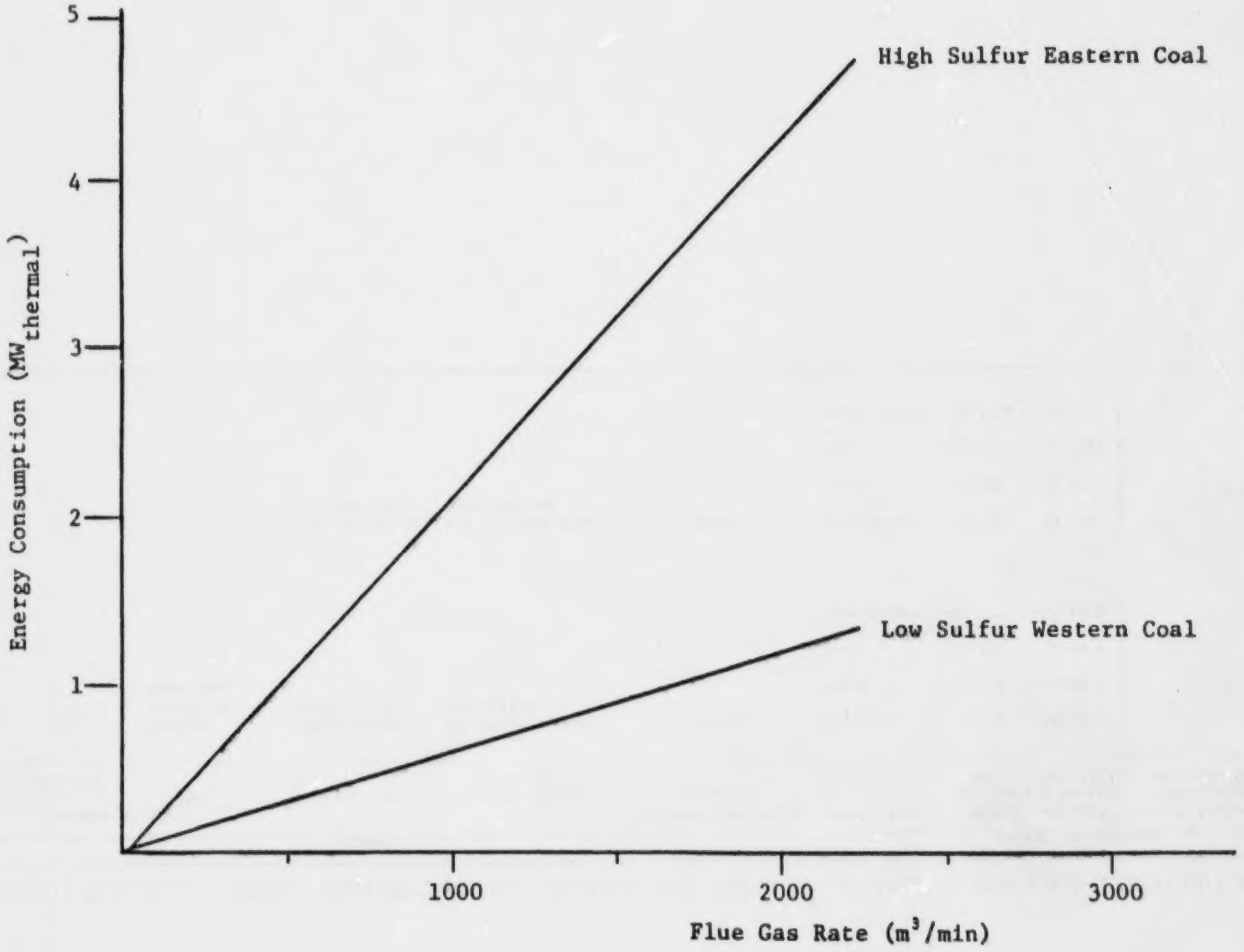


Figure 5.2.2-1. Energy consumption of parallel flow SCR $NO_x/SO_x$ FGT systems for coal-fired boilers.

The energy impact of FGT controls applied to modified or reconstructed facilities (retrofit application) will be the same or greater than that for new facilities. If flue gas can be taken from the economizer of the existing boiler at 350-400°C and returned upstream of any existing heat exchange equipment, then there will be no additional energy impact.

If the flue gas is only available at a lower temperature ( 350-400°C) then a heater will be required. The energy impact of the heater will depend on the temperature of the flue gas. The energy used in heating the gas is not completely lost since the air preheater can recover about 85% of the energy supplied by the gas heater. The heater will probably be oil-fired for ease of control.

These results indicate that, on retrofit applications, there is an energy incentive to obtain the flue gas at the necessary reaction temperature in order to avoid flue gas heating. Other energy impacts would be the same as those for new facilities.

### 5.2.3 Energy Impact of $NO_x/SO_x$ Controls for Oil-Fired Boilers

Three types of energy are utilized by these systems--electrical, steam and fuel. The amounts of each type are shown in Table 5.2.3-1. Energy consumption of each energy type is of the same order of magnitude. Also shown is the heat credit that is obtained by partially recovering heat from the energy inputs with the air preheater.

Since only one case is considered, the variables that affect the energy impact cannot be quantified. It can be stated qualitatively, however, that the primary variables that affect energy usage are oil sulfur content and control level (both $NO_x$ and $SO_x$). The effect of fuel sulfur content was examined in the section on coal-fired applications.

TABLE 5.2.3-1. ENERGY CONSUMPTION FOR $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR OIL FIRED BOILERS

| System | | | | | | Energy Consumption | | | |
|---|---|---|---|---|---|---|---|---|---|
| Standard Boiler | | | Oil Type | Type and Level of Control | Control Efficiency (% $NO_x/SO_x$) | Energy Types | Energy Consumed By Control Device | | % Increase in Energy Use Over Uncontrolled Boiler |
| Heat Input | | Type | | | | | | | |
| $MW_t$ | (MBtu/hr) | | | | | | $MW_t$ | (MBtu/hr) | |
| 44 | 150 | Package Watertube | Residual | SCR Parallel Flow Intermediate | 80/85 | Electrical | 1.26 | (4.29) | 4.4 |
| | | | | | | Steam | 1.16 | (3.96) | |
| | | | | | | Fuel | 2.16 | (7.37) | |
| | | | | | | Heat Credit | -(2.63) | -(8.98) | |

The energy impact of FGT controls applied to modified or reconstructed facilities (retrofit application) will be the same or greater than that for new facilities. If flue gas can be taken from the economizer of the existing boiler at 350-400°C and returned upstream of any existing heat exchange equipment, then there will be no additional energy impact.

If the flue gas is only available at a lower temperature (<350-400°C) then a heater will be required. The energy impact of the heater will depend on the temperature of the flue gas. The energy used in heating the gas is not completely lost since the air preheater can recover about 85% of the energy supplied by the heater. The heater will probably be oil-fired for ease of control.

These results indicate that, on retrofit applications, there is an energy incentive to obtain the flue gas at the necessary reaction temperature in order to avoid flue gas heating. Other energy impacts would be the same as those for new facilities.

## 5.3 SUMMARY

All of the $NO_x$-only systems and cases required <1% of the total heat input to the boiler. By far, the item contributing the most to energy consumption was the flue gas fan which supplied the draft loss caused by the catalyst bed.

The parallel flow systems appear to use more energy than the moving bed systems; however, both are of the same order of magnitude. Within the accuracy of the calculations, the systems examined should be considered to have approximately similar energy impacts.

The $NO_x/SO_x$ systems require 2-8% of the total heat input to the boiler. This is primarily due to the requirement for $SO_2$ workup. Although this requirement is higher than that for $NO_x$-only processes, it may be less than that for the combination of $NO_x$-only and FGD.

# REFERENCES

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# SECTION 6

# ENVIRONMENTAL IMPACT OF CANDIDATES FOR BEST EMISSION CONTROL SYSTEMS

## 6.1 INTRODUCTION

The three best candidate control techniques identified in Section III are as follows:

SCR - Fixed Packed Bed
SCR - Parallel Flow
SCR - Moving Bed

These techniques have similar environmental concerns as they all utilize $NH_3$ to reduce $NO_x$ to $N_2$. All are capable of achieving the levels of control considered in this study, although the applicability of a particular system is fuel dependent. While $NO_x$ reduction is the primary beneficial environmental impact of these systems, particulate removal is a secondary beneficial impact of the moving bed systems. The moving bed system will reduce the level of particulates in the flue gas by 70-80%.[1,2] The particulates are embedded on the catalyst (rings or granules) as the catalyst moves downward through the reactor. The dirty catalyst is removed to a vibrating screen which separates the dust and the clean catalyst is then recycled to the top of the reactor. Pilot plant tests on the moving bed reactor have shown it capable of handling <1 $g/Nm^3$ of particulates.[3] The uncontrolled particulate levels in the flue gas from the pulverized coal (5-9 $g/Nm^3$), spreader stoker (3.5-6.3 $g/nM^3$), chaingrate and underfeed (both 1.4-2.4 $g/Nm^3$) standard boilers are all greater than this figure. As a result, moving bed systems are not considered for application to the coal-fired standard boilers. The fixed packed bed system cannot tolerate particulates so it is applied only to natural gas- and distillate oil-fired flue gas which have low particulate loadings (13 and 19 mg/Nm$^3$, respectively, for the standard boilers). Conversely, the parallel flow system can tolerate full particulate loadings (up to 20 g/nM$^3$)[4] as the open passageways present unobstructed paths for particulates to pass through with little impaction on the catalyst surface.

There are some potential adverse environmental impacts of the selective catalytic reduction processes. First, the use of $NH_3$ as the gaseous reducing agent introduces the possibility of ammonia emissions. Commercial operations of the three reactor types on industrial and utility boilers have demonstrated emissions of <10 ppm $NH_3$ at the $NH_3$:$NO_x$ mole ratio required for stringent level of control. These levels are shown graphically in Figures 6.1-1 through 6.1-3 on the following pages. (These plots are composites of the available commercial data.) The only data available on $NH_3$ emissions are from Japanese process vendors and these data indicate $NH_3$ emissions to be <10 ppm. This number may be optimistic, especially considering that currently there is no continuous monitoring technique for measuring $NH_3$ in the presence of $SO_x$. The data, therefore, represent spot measurements and not continuous data. It seems reasonable to assume that 10 ppm represents a minimum level of $NH_3$ emissions.

A potential environmental problem in commercial SCR operations is the formation of ammonium bisulfate, $NH_4HSO_4$, or ammonium sulfate, $(NH_4)_2SO_4$. The presence of $NH_3$, $SO_3$, and $H_2O$ in the hot flue gas leads to the formation of liquid $NH_4HSO_4$ upon cooling to approximately 180-220°C by the following reaction.

$$NH_3(g) + SO_3(g) + H_2O(g) \rightleftarrows NH_4HSO_4(\ell) \qquad (6\text{-}1)$$

This can create a plugging and corrosion problem in heat exchange equipment, particularly for boilers burning medium- or high-sulfur fuels. Further cooling to about 190°C precipitates the formation of solid ammonium sulfate [$(NH_4)_2SO_4$] by the following reaction.

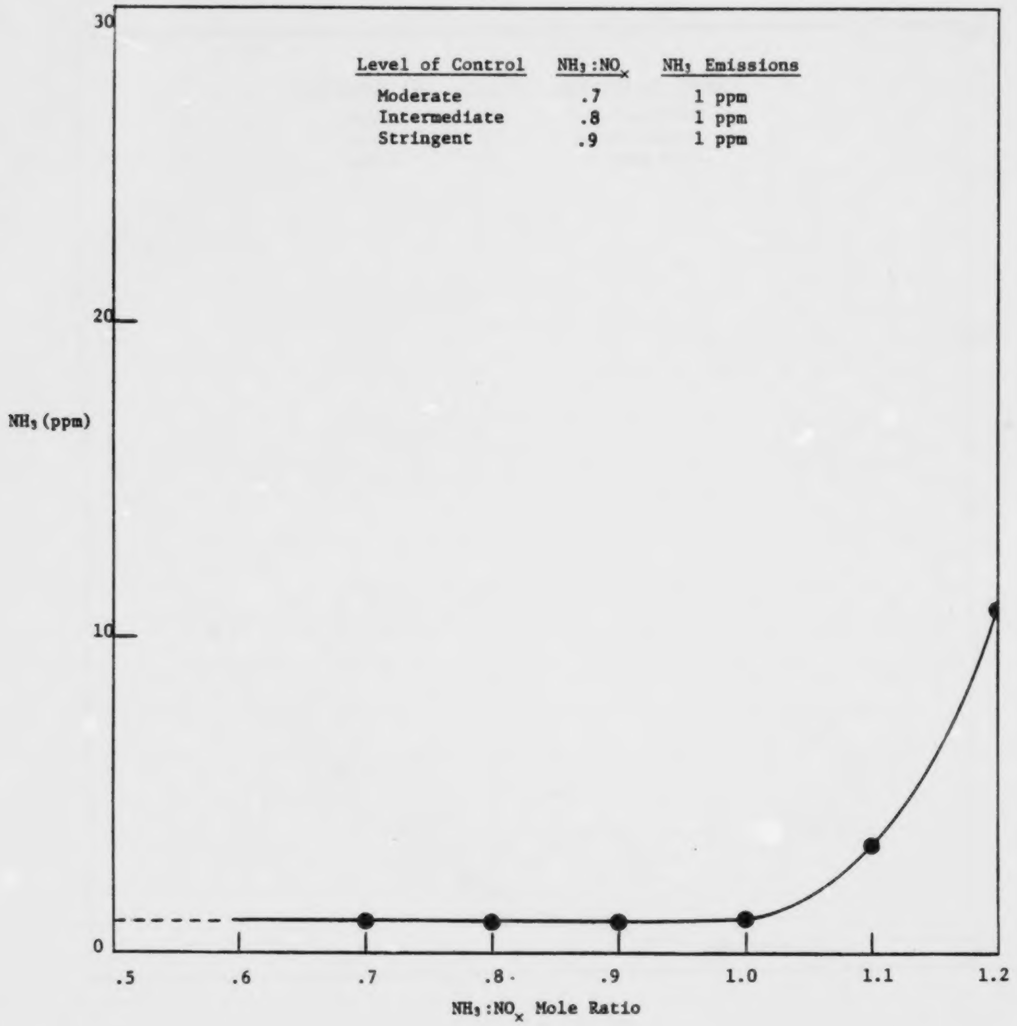
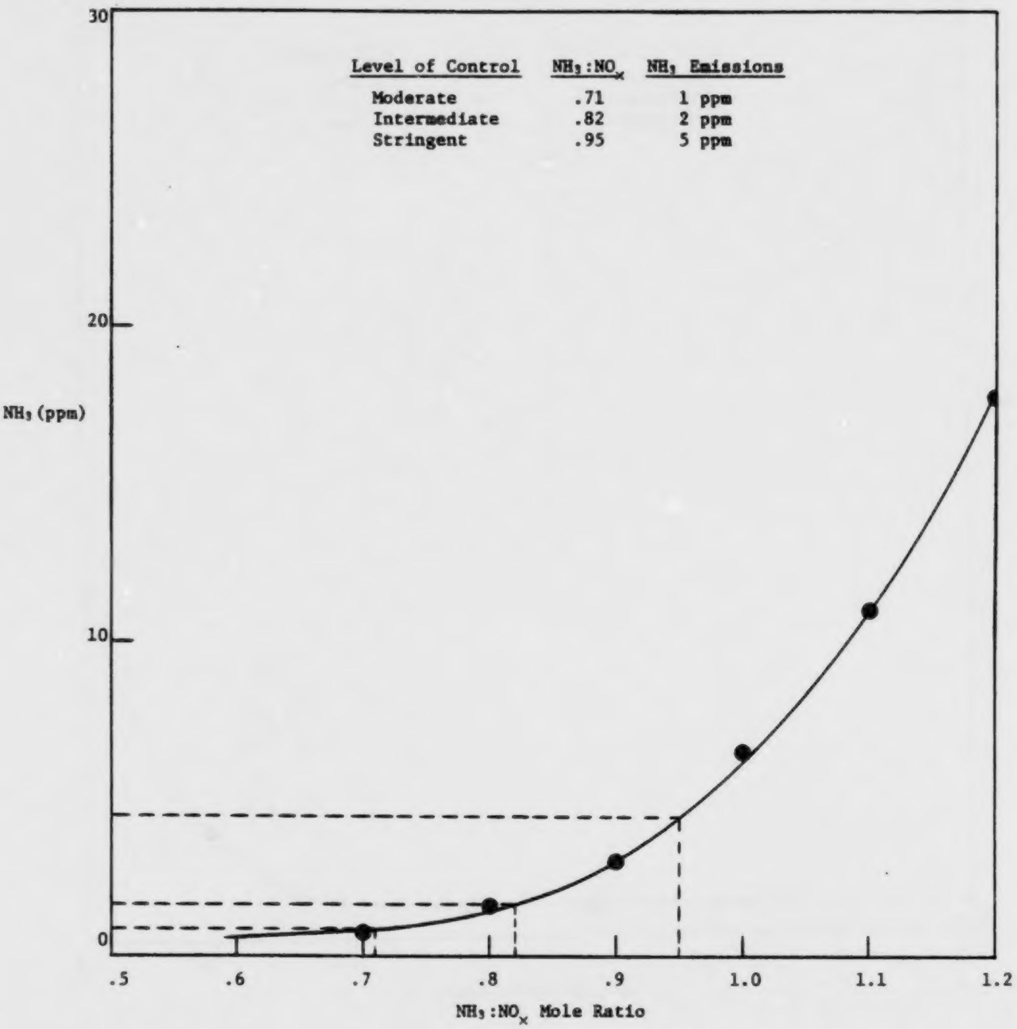


Figure 6.1-1. $NH_3$ Emissions - Fixed Packed Bed Reactor.[5,6,7]



| Level of Control | $NH_3:NO_x$ | $NH_3$ Emissions |
|---|---|---|
| Moderate | .71 | 1 ppm |
| Intermediate | .82 | 2 ppm |
| Stringent | .95 | 5 ppm |

Figure 6.1-2. $NH_3$ Emissions - Parallel Flow Reactor.[8,9,10,11,12]

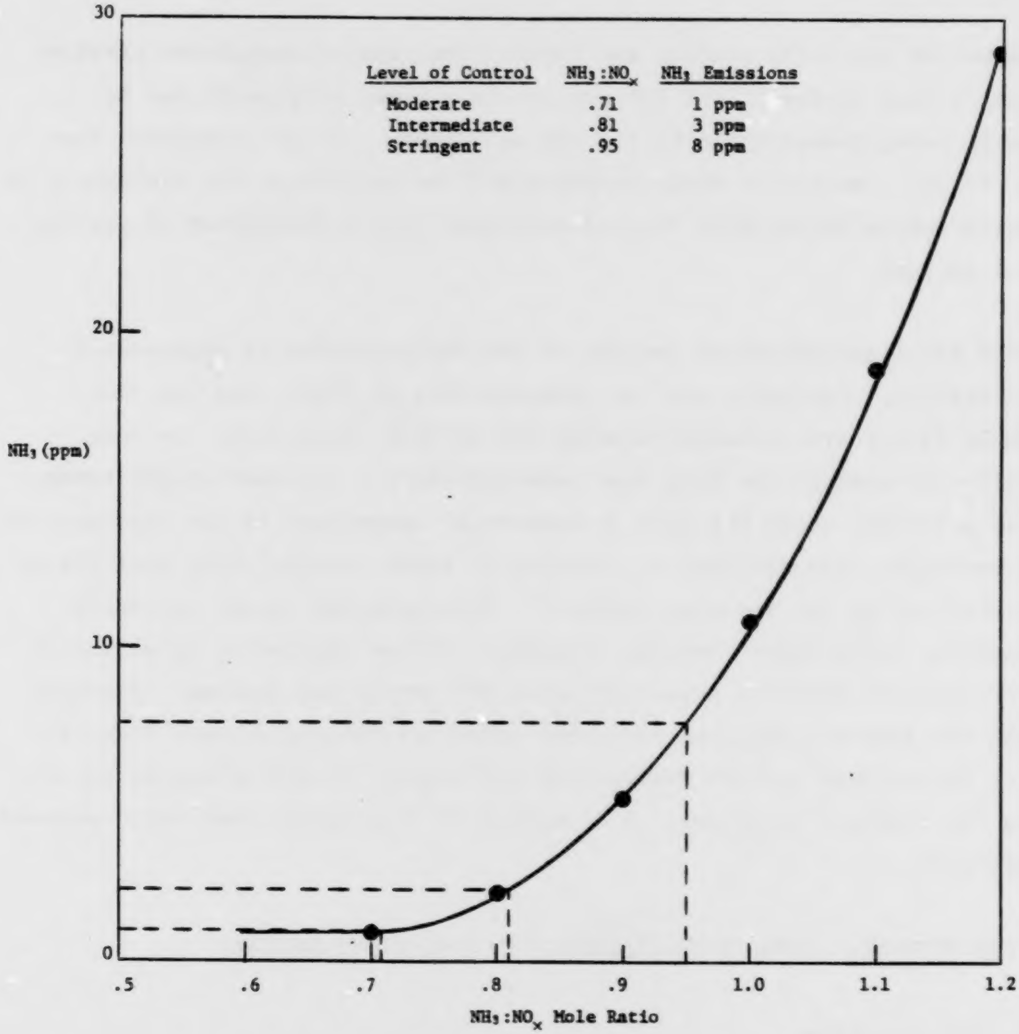


Figure 6.1-3. $NH_3$ Emissions - Moving Bed Reactor.[13,14,15,16,17]

$$NH_4HSO_4(\ell) + NH_3(g) \rightleftarrows (NH_4)_2SO_4(s) \qquad (6\text{-}2)$$

The impact of the solid sulfate and liquid bisulfate on downstream particulate collection equipment and FGD systems is unknown at present and is currently being investigated by the EPA and others. It is speculated that minor, if any, amounts of these sulfates will be emitted to the atmosphere in situations where particulate control equipment exists downstream of the $NO_x$ control system.

The final environmental concern of the SCR processes is disposed of spent catalyst. Catalysts such as titanium dioxide ($TiO_2$) and vanadium pentoxide ($V_2O_5$) are probably recycled due to their high cost. To date, virtually no catalyst has been used commercially yet for over 10,000 hours, and, as a result, there has been no commercial experience on the treatment of spent catalyst. Reprocessing or disposal of spent catalyst will most likely be carried out by the catalyst vendor.[27] This question is not currently addressed in literature. Another potential problem related to catalysts is that of catalyst erosion, especially with the moving bed systems. Catalyst erosion may generate small particulates which may present a stack fume problem if particulate control devices are not present or not effective at removing the catalyst particles. No problems of this nature have been reported at this time.

## 6.2 ENVIRONMENTAL IMPACTS OF CONTROLS FOR COAL-FIRED BOILERS

### 6.2.1 Air Pollution

The emission rates for primary and secondary pollutants are presented in Tables 6.2.1-1 through 6.2.1-12 on the following pages. There are three tables for each of the 4 coal-fired standard boilers. Each table is broken down according to coal type (high sulfur eastern and low sulfur western) and control level (uncontrolled, moderate, intermediate, and stringent). For each entry the impact on the primary pollutant, $NO_x$, is shown. Then, the adverse impact of the secondary pollutant, $NH_3$, is given for each case.

TABLE 6.2.1-1. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Pulverized Coal
Heat Rate: 200 MBtu/hr
Coal: High Sulfur Eastern

| | | $NO_x$ | | Particulates | | $NH_3$ | | |
|---|---|---|---|---|---|---|---|---|
| Control level | Type of control | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | Bisulfate |
| Uncontrolled | - | 19.2<br>(152.46) | 328.0<br>(.762) | 181.2<br>(1436.5) | 3090<br>(7.18) | 0 | 0 | 0 |
| Moderate-70% | SCR-Parallel Flow | 5.77<br>(45.7) | 98.2<br>(.229) | Negligible Effect | | .0154<br>(.122) | .261<br>(.000608) | See Text |
| Stringent-90% | SCR-Parallel Flow | 1.92<br>(15.2) | 32.8<br>(.0762) | Negligible Effect | | .0767<br>(.608) | 1.31<br>(.00304) | See Text |

TABLE 6.2.1-2. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Pulverized Coal
Heat Rate: 200 MBtu/hr
Coal: Low Sulfur Western

| | | $NO_x$ | | Particulates | | $NH_3$ | | |
|---|---|---|---|---|---|---|---|---|
| Control level | Type of control | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | Bisulfate |
| Uncontrolled | - | 23.7<br>(187.56) | 403.0<br>(.938) | 113.5<br>(900.3) | 1936<br>(4.50) | 0 | 0 | 0 |
| Moderate-70% | SCR-Parallel Flow | 7.10<br>(56.3) | 121.0<br>(.282) | Negligible Effect | | .0159<br>(.126) | .272<br>(.000632) | See Text |
| Stringent-90% | SCR-Parallel Flow | 2.37<br>(18.8) | 40.3<br>(.0938) | Negligible Effect | | .0797<br>(.632) | 1.36<br>(.00316) | See Text |

TABLE 6.2.1-3. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Spreader Stoker
Heat Rate: 150 MBtu/hr
Coal: High Sulfur Eastern

| Control level | Type of control | $NO_x$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 12.0<br>(95.4) | 273.0<br>(.636) | 111.0<br>(876.4) | 2512<br>(5.84) | 0 | 0 | 0 |
| Intermediate-80% | SCR-Parallel Flow | 2.41<br>(19.1) | 54.7<br>(.127) | Negligible Effect | | .0266<br>(.211) | .604<br>(.00140) | See Text |

TABLE 6.2.1-4. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Spreader Stoker
Heat Rate: 150 MBtu/hr
Coal: Low Sulfur Western

| Control level | Type of control | $NO_x$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 14.8<br>(117.15) | 336.0<br>(.781) | 69.1<br>(548.3) | 1572<br>(3.66) | 0 | 0 | 0 |
| Intermediate-80% | SCR-Parallel Flow | 2.95<br>(23.4) | 67.2<br>(.156) | Negligible Effect | | .0273<br>(.217) | .622<br>(.00145) | See Text |

TABLE 6.2.1-5. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Chaingrate
Heat Rate: 75 MBtu/hr
Coal: High Sulfur Eastern

| Control level | Type of control | $NO_x$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 6.02<br>(47.7) | 273.0<br>(.636) | 21.2<br>(168.5) | 966.0<br>(2.25) | 0 | 0 | 0 |
| Moderate-70% | SCR-Parallel Flow | 1.80<br>(14.3) | 82.0<br>(.191) | Negligible Effect | | .00662<br>(.0525) | .301<br>(.000700) | See Text |
| Intermediate-80% | SCR-Parallel Flow | 1.20<br>(9.54) | 54.7<br>(.127) | Negligible Effect | | .0132<br>(.105) | .602<br>(.00140) | See Text |
| Stringent-90% | SCR-Parallel Flow | .602<br>(4.77) | 27.3<br>(.0636) | Negligible Effect | | .0331<br>(.262) | 1.50<br>(.00350) | See Text |

TABLE 6.2.1-6. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Chaingrate
Heat Rate: 75 MBtu/hr
Coal: Low Sulfur Western

| Control level | Type of control | $NO_x$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 7.40<br>(58.65) | 336.0<br>(.782) | 13.3<br>(105.6) | 605.0<br>(1.41) | 0 | 0 | 0 |
| Moderate-70% | SCR-Parallel Flow | 2.22<br>(17.6) | 101.0<br>(.235) | Negligible Effect | | .00683<br>(.0542) | .311<br>(.000723) | See Text |
| Intermediate-80% | SCR-Parallel Flow | 1.48<br>(11.7) | 67.2<br>(.156) | Negligible Effect | | .0137<br>(.108) | .621<br>(.00145) | See Text |
| Stringent-90% | SCR-Parallel Flow | .740<br>(5.87) | 33.6<br>(.0782) | Negligible Effect | | .0342<br>(.271) | 1.55<br>(.00361) | See Text |

TABLE 6.2.1-7. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Underfeed Stoker
Heat Rate: 30 MBtu/hr
Coal: High Sulfur Eastern

| Control level | Type of control | $NO_x$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 2.40<br>(19.05) | 273.0<br>(.635) | 8.49<br>(67.31) | 965.0<br>(2.24) | 0 | 0 | 0 |
| Intermediate-80% | SCR-Parallel Flow | .480<br>(3.81) | 54.6<br>(.127) | Negligible Effect | | .00529<br>(.0419) | .601<br>(.00140) | See Text |

TABLE 6.2.1-8. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS[1,2,8-17]

Standard Boiler: Underfeed Stoker
Heat Rate: 30 MBtu/hr
Coal: Low Sulfur Western

| Control level | Type of control | $NO_x$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 2.95<br>(23.40) | 335.0<br>(.780) | 5.31<br>(42.12) | 604.0<br>(1.40) | 0 | 0 | 0 |
| Intermediate-80% | SCR-Parallel Flow | .590<br>(4.68) | 67.1<br>(.156) | Negligible Effect | | .00544<br>(.0431) | .618<br>(.00144) | See Text |

## TABLE 6.2.1-9. AIR POLLUTION IMPACTS FROM BEST $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Standard Boiler: Pulverized Coal
Heat Rate: 200 MBtu/hr
Coal: High Sulfur Eastern

| Control level | Type of control | $NO_x$ | | $SO_2$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 19.2<br>(152.5) | 328.0<br>(0.762) | 142.0<br>(1127.0) | 2423.0<br>(5.64) | 181.0<br>(1437.0) | 3090.0<br>(7.18) | 0 | 0 | 0 |
| Intermediate<br>(80% $NO_x$)<br>(85% $SO_2$) | SRC-Parallel Flow | 3.85<br>(30.5) | 65.6<br>(0.153) | 21.3<br>(169.0) | 363.0<br>(0.865) | Negligible Effect | | 0.307<br>(2.43) | 5.22<br>(0.0122) | See Text |

## TABLE 6.2.1-10. AIR POLLUTION IMPACTS FROM BEST $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Standard Boiler: Pulverized Coal
Heat Rate: 200 MBtu/hr
Coal: Low Sulfur Western

| Control level | Type of control | $NO_x$ | | $SO_2$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 23.7<br>(187.6) | 403.0<br>(0.938) | 30.0<br>(237.6) | 511.0<br>(1.19) | 113.5<br>(900.3) | 1936.0<br>(4.50) | 0 | 0 | 0 |
| Intermediate<br>(80% $NO_x$)<br>(85% $SO_2$) | SRC-Parallel Flow | 4.73<br>(37.5) | 80.6<br>(0.188) | 4.49<br>(35.6) | 7.65<br>(0.178) | Negligible Effect | | 0.318<br>(2.52) | 5.42<br>(0.0126) | See Text |

TABLE 6.2.1-11. AIR POLLUTION IMPACTS FROM BEST $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Standard Boiler: Underfeed Stoker
Heat Rate: 30 MBtu/hr
Coal: High Sulfur Eastern

| | | $NO_x$ | | $SO_2$ | | Particulates | | $NH_3$ | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Control level | Type of control | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | Bisulfate |
| Uncontrolled | - | 2.40<br>(19.05) | 273.0<br>(0.635) | 21.3<br>(168.9) | 2421.0<br>(5.63) | 8.49<br>(67.31) | 965.0<br>(2.24) | 0 | 0 | 0 |
| Intermediate<br>(80% $NO_x$)<br>(85% $SO_2$) | SRC-Parallel Flow | 0.481<br>(3.81) | 54.6<br>(0.127) | 3.20<br>(25.3) | 363.0<br>(0.845) | Negligible Effect | | 0.0529<br>(0.419) | 6.01<br>(0.0140) | See Text |

TABLE 6.2.1-12. AIR POLLUTION IMPACTS FROM BEST $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR COAL-FIRED BOILERS

Standard Boiler: Underfeed Stoker
Heat Rate: 30 MBtu/hr
Coal: Low Sulfur Western

| | | $NO_x$ | | $SO_2$ | | Particulates | | $NH_3$ | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Control level | Type of control | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | Bisulfate |
| Uncontrolled | - | 2.95<br>(23.40) | 335.0<br>(0.780) | 4.49<br>(35.6) | 510.0<br>(1.19) | 5.31<br>(42.12) | 604.0<br>(1.40) | 0 | 0 | 0 |
| Intermediate<br>(80% $NO_x$)<br>(85% $SO_2$) | SRC-Parallel Flow | 0.590<br>(4.58) | 67.1<br>(0.156) | 0.674<br>(5.34) | 76.8<br>(0.179) | Negligible Effect | | 0.0544<br>(0.431) | 6.18<br>(0.0144) | See Text |

Table 6.2.1-13 shows the uncontrolled $NO_x$ emissions for each standard boiler and the SIP level for each fuel. One can see that a majority of the uncontrolled emissions are less than SIP allowable levels. For coal, the worst case is the pulverized coal boiler burning low sulfur western coal. To meet the SIP control level the degree of removal required is as follows.

$$\% \text{ reduction} = \frac{.938 - .7}{.938} \times 100\% = 25\%$$

This level of control is easily obtained via combustion modifications,[18] therefore, this study does not address the SIP control level.

Also, in Tables 6.2.1-1 through 6.2.1-12 the secondary pollutant bisulfate is not quantified, for several reasons. First, kinetic and thermodynamic data for the reaction

$$NH_3(g) + SO_3(g) + H_2O(g) \rightleftarrows NH_4HSO_4(\ell)$$

have not yet been evaluated. Therefore, the extent of reaction cannot be determined for the residence time of the flue gas in the duct. Second, bisulfate emissions are not constant since they are at a high level during soot blowing and at a lower level during other periods. Finally, it is not known how much of the bisulfate is caught by downstream particulate removal equipment (assuming that the equipment is present) and how much is emitted. A hypothetical calculation can be made for the case of the pulverized coal standard boiler burning high sulfur eastern coal (Table 6.2.1-1) equipped with an SCR parallel flow control system operating at the stringent level of control. The amount of bisulfate formed is as follows.

## TABLE 6.2.1-13. $NO_x$ EMISSION LEVELS AND SIP CONTROL LEVELS
(All values in $lb/10^6$ Btu)

| | SIP control | Uncontrolled standard boilers | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| | | Natural gas | Distillate oil | Residual oil | Coal type | Underfeed stoker | Chaingrate | Spreader stoker | Pulverized coal |
| Gas | .2 | .175 | | | | | | | |
| Oil | .3 | | .159 | .400 | | | | | |
| Coal | .7 | | | | High S Eastern | .635 | .636 | .636 | .762 |
| | | | | | Low S Eastern | .545 | .544 | .543 | .653 |
| | | | | | Low S Western | .780 | .782 | .781 | .938 |

Assuming all $NH_3 \rightarrow NH_4HSO_4$:

$$\text{wt. bisulfate} = \left(.0767 \frac{\text{g } NH_3}{\text{s}}\right)\left(\frac{\text{g-mole } NH_3}{17.0\text{g } NH_3}\right)\left(\frac{\text{g-mole } NH_4HSO_4}{\text{g-mole } NH_3}\right)\left(\frac{115\text{g } NH_4HSO_4}{\text{g-mole } NH_4HSO_4}\right)$$

$$= .519 \frac{\text{g } NH_4HSO_4}{\text{s}} = \left(19 \frac{\text{tons}}{\text{year}}\right)$$

There are beneficial aspects of this reaction. The bisulfate formation ties up $SO_3$ which is more hazardous than $SO_2$ and is difficult to catch with FGD.[19] If the bisulfate can be collected adequately and disposed of safely, an overall improvement could be achieved.

The fate of bisulfate formed in this manner has not been completely resolved and is currently an aspect of $NO_x$ control receiving much attention. In Japan there have been problems at some installations of precipitation of the bisulfate or sulfate on elements of regenerative air preheaters and tubes of tubular air preheaters. This deposit can be removed by periodically water-washing the air preheater.[20] Bisulfate or sulfate particulates that pass through the air preheater may be collected by downstream particulate control or FGD equipment if such equipment exists. The collection efficiency of particulate control equipment cannot be determined quantitatively without knowing the particle size and resistivity (for ESP's) or the K-factor and concentration (for baghouses). These data are not currently known for these compounds, however, it is reasonable to expect that some fraction of the particles will be collected.[21] Similarly, the collection efficiency for an FGD unit has not been examined. Where neither particulate control nor FGD equipment exist, there may be stack emissions of sulfates. An FGD system may also absorb $NH_3$ emitted by an SCR system, however, the removal cannot be determined from the information currently available.

To reduce the adverse environmental effects, improved combustion control utilizing less $O_2$ minimizes the formation of $NO_x$. It also minimizes formation of $SO_3$ which is necessary for ammonium bisulfate formation. Careful operation of the FGT system should keep the $NH_3$ injection ratio as low as possible to minimize $NH_3$ emissions and bisulfate formation. Also, careful operation of downstream heat exchange equipment to keep the flue gas above the acid ($SO_3$) dew point is required. The use of corrosion-resistant material in any heat exchanger is advisable where $NH_4HSO_4$ deposits are probable.[22] A multitube type heat exchanger with the tubes placed vertically is a possible configuration to prevent bisulfate deposits from causing problems. Any bisulfate liquid formed in the tube will drip downward as long as the temperature of the tube is kept above the melting point of bisulfate.[23] It will be necessary to design the exchanger out of corrosion resistant materials.

### 6.2.2 Water Pollution

There are no water streams that are associated with $NO_x$-only SCR systems, however, there is one potential source of water pollution. In some Japanese installations $NH_4HSO_4$ deposits are removed from the air preheater by water washing. The blowdown from this operation will contain both ammonium and sulfate ions which, if not treated, present a water pollution source. Since the amounts of $NH_4HSO_4$ and water are not known it is possible to estimate the concentration or flow additional of this potential source.

### 6.2.3 Solid Waste

The only potential solid waste associated with this system is spent catalyst. Presently, the life of SCR catalysts is 1-2 years and the topic of recycling is not addressed in the literature. Since, to date, virtually no commercial SCR units have operated long enough to require catalyst replacement, there is no commercial experience on the treatment of spent catalyst.[27] With the high cost of some of the catalysts, recycling seems to be warranted. Recycling is feasible where the spent catalyst support is composed of valuable materials such as titanium, stainless steel, or possibly a ceramic material. Alumina catalyst supports probably do not warrant recycling unless required for environmental reasons. These catalysts can probably be disposed of in the same manner as other industrial cataylsts.

The amount of catalyst that must be recycled or disposed of is one reactor volume since replacement involves total catalyst replacement. The actual frequency of catalyst replacement is unique to each specific process, however, catalyst lifetimes are typically one or more years.

Few process vendors have published their catalyst formulations since the field is very competitive at the present time. Base metal oxides are typically used.[24] The environmental impact of catalyst disposal will depend on what materials and compounds are involved.

### 6.2.4 Other Environmental Impacts

The only other environmental impact is a secondary impact. $NH_3$ is commonly made from natural gas and its consumption is considered a secondary environmental impact. This impact is quantified in a report on the impact of $NO_x$ regulations on the $NH_3$ industry. The report was prepared by TVA under contract to EPA-IERL. Other adverse environmental impacts (noise, thermal pollution, electrical discharges, etc.) are not present with SCR systems.

### 6.2.5 Environmental Impact on Modified and Reconstructed Facilities

The environmental impacts of a new facility and a retrofitted facility should be similar. There is not enough difference between new and retrofit systems to indicate that environmental impacts would be significantly different with retrofitted systems.

## 6.3 ENVIRONMENTAL IMPACTS OF CONTROLS FOR OIL-FIRED BOILERS

### 6.3.1 Air Pollution

Emission rates for primary and secondary pollutants are presented in Tables 6.3.1-1 and 6.3.1-2 on the next pages. The tables are organized by fuel type (residual and distillate oil); control level (uncontrolled, moderate, and stringent); and type of control (residual - SCR moving bed and SCR parallel flow; distillate - SCR fixed packed bed). The impact on the primary pollutant, $NO_x$, is given for each case. Also, the moving bed's impact on particulates of residual oil is shown. The adverse impact of the secondary pollutant, $NH_3$, is given for each entry.

Table 6.2.1-13 shows that only the residual oil-fired flue gas has uncontrolled $NO_x$ emissions greater than the SIP control level. To achieve the SIP level of control the removal required is as follows.

$$\begin{aligned} \%\ \text{reduction} &= \frac{0.4-0.3}{0.4} \times 100\% \\ &= 25\% \end{aligned}$$

This control level is readily achieved by combustion modifications; hence, FGT to achieve the SIP control level is not examined for oil-fired boilers.

In Table 6.3.1-1 the secondary pollutant bisulfate is not quantified. This is due to a lack of developed kinetic and thermodynamic data to predict the extent of reaction. Also, removal levels are not constant since the degree of downstream particulate removal is uncertain. However, one can see that the bisulfate problem is worse for residual oil than for distillate oil because there is more $SO_3$ available for reaction. Bisulfate is formed by a one-to-one reaction between $NH_3$, $SO_3$ and $H_2O$.

$$NH_3(g) + SO_3(g) + H_2O(g) \rightleftarrows NH_4HSO_4(\ell)$$

## TABLE 6.3.1-1. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR OIL-FIRED BOILERS[1,2,5-17]

| Standard boiler | | | | $NO_x$ | | Particulates | | $NH_3$ | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Heat rate (MBtu/hr) | Type | Control level | Type of control | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | Bisulfate |
| 150 | Residual | Uncontrolled | - | 7.57<br>(60.0) | 172.0<br>(.400) | 4.16<br>(33.0) | 94.6<br>(.220) | 0 | 0 | 0 |
| | | Moderate - 70% | SCR Moving Bed | 2.27<br>(18.0) | 51.6<br>(.120) | 1.25<br>(9.90) | 28.4<br>(.0660) | .00957<br>(.0759) | .218<br>(.000506) | See Text |
| | | Stringent - 90% | SCR Moving Bed | .757<br>(6.00) | 17.2<br>(.0400) | 1.25<br>(9.90) | 28.4<br>(.0660) | .0766<br>(.607) | 1.74<br>(.00405) | See Text |
| | | Moderate - 70% | SCR Parallel Flow | 2.27<br>(18.0) | 51.6<br>(.120) | Negligible Effect | | .00957<br>(.0759) | .218<br>(.000506) | See Text |
| | | Stringent - 90% | SCR Parallel Flow | .757<br>(6.00) | 17.2<br>(.0400) | Negligible Effect | | .0479<br>(.380) | 1.09<br>(.00253) | See Text |
| 30 | Residual | Uncontrolled | - | 2.02<br>(16.0) | 229.0<br>(0.533) | .580<br>(4.60) | 65.9<br>(.153) | 0 | 0 | 0 |
| | | Moderate - 70% | SCR Moving Bed | 0.606<br>(4.80) | 68.7<br>(0.160) | .174<br>(1.38) | 20.1<br>(.0459) | .00201<br>(.0160) | .229<br>(.000533) | See Text |
| | | Stringent - 90% | SCR Moving Bed | 0.202<br>(1.60) | 22.9<br>(0.0533) | .174<br>(1.38) | 20.1<br>(.0459) | .0161<br>(.128) | 1.83<br>(.00426) | See Text |
| | | Moderate - 70% | SCR Parallel Flow | 0.606<br>(4.80) | 68.7<br>(0.160) | Negligible Effect | | .00201<br>(.0160) | .229<br>(.000533) | See Text |
| | | Stringent - 90% | SCR Parallel Flow | 0.202<br>(1.60) | 22.9<br>(0.0533) | Negligible Effect | | .0101<br>(.0798) | 1.14<br>(.00266) | See Text |
| 75 | Distillate | Uncontrolled | - | 2.99<br>(23.76) | 68.0<br>(0.158) | 1.02<br>(8.10) | 46.4<br>(.108) | 0 | 0 | 0 |
| | | Moderate - 70% | SCR Fixed Packed Bed | 0.898<br>(7.13) | 20.4<br>(0.047) | Negligible Effect | | .00502<br>(.0398) | .228<br>(.000531) | Less than Residual |
| | | Stringent - 90% | SCR Fixed Packed Bed | 0.299<br>(2.38) | 6.80<br>(0.0158) | Negligible Effect | | .00502<br>(.0398) | .228<br>(.000531) | Less than Residual |
| 15 | Distillate | Uncontrolled | - | .300<br>(2.38) | 68.2<br>(.159) | Negligible Amount | | 0 | 0 | 0 |
| | | Moderate - 70% | SCR Fixed Packed Bed | .0900<br>(.714) | 20.5<br>(.0476) | | | .00109<br>(.00864) | .248<br>(.000576) | Less than Residual |
| | | Stringent - 90% | SCR Fixed Packed Bed | .0300<br>(.238) | 6.82<br>(.0159) | | | .00109<br>(.00864) | .248<br>(.000576) | Less than Residual |

### TABLE 6.3.1-2. AIR POLLUTION IMPACTS FROM BEST $NO_x/SO_x$ FGT CONTROL TECHNIQUES FOR OIL-FIRED BOILERS

Boiler Type: Watertube
Heat Rate: 150 MBtu/hr
Oil: Residual

| Control level | Type of control | $NO_x$ | | $SO_2$ | | Particulates | | $NH_3$ | | Bisulfate |
|---|---|---|---|---|---|---|---|---|---|---|
| | | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) | |
| Uncontrolled | - | 7.57<br>(60.0) | 172.0<br>(0.400) | 59.4<br>(471.0) | 1350.0<br>(3.14) | 4.16<br>(33.0) | 94.6<br>(0.220) | 0 | 0 | 0 |
| Intermediate<br>(80% $NO_x$)<br>(85% $SO_2$) | SRC-Parallel Flow | 1.51<br>(12.0) | 34.4<br>(0.0800) | 8.91<br>(70.7) | 203.0<br>(0.471) | Negligible Effect | | 0.191<br>(1.52) | 4.36<br>(0.0101) | See Text |

The flue gas $SO_3$ concentration can be calculated as follows:

Residual Oil

Flue gas $= 46{,}700\ \frac{ft^3}{min}\left(\frac{32 + 460\ °R}{400 + 460\ °R}\right)\left(\frac{lb\text{-}mole}{359\ scf}\right)\left(\frac{60\ min}{hr}\right)$

$= 4465$ lb-moles/hr

$SO_2$ $= 471.0\ \frac{lbs}{hr}\left(\frac{lb\text{-}mole}{64.0\ lb}\right)$

$= 7.359$ lb-moles/hr

Fuel S = 3.0%

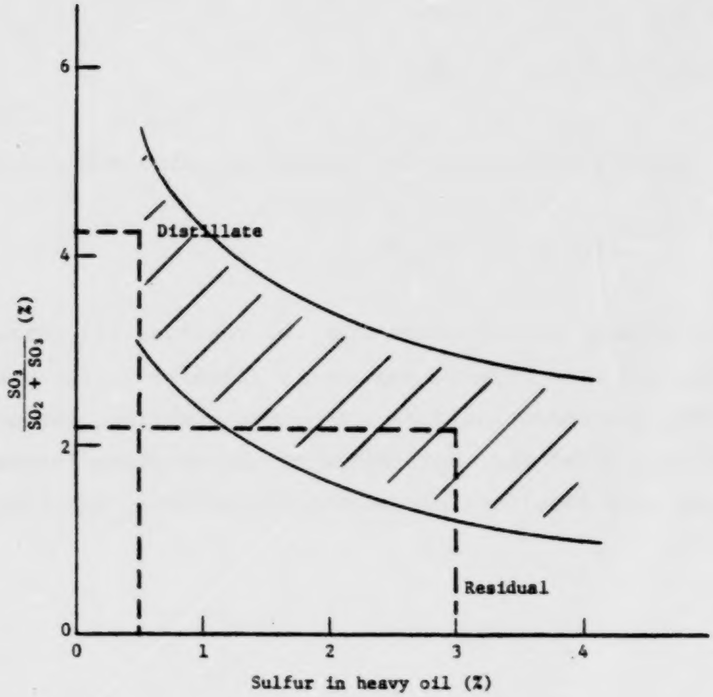


Figure 6.3.1-1. Formation ratio of $SO_3$.[25]

from Figure 6.3.1-1 % $SO_3$ = 2.3%

The $SO_3$ concentration can be determined by calculation to be

$$[SO_3] \quad = 39 \text{ ppm}$$

Distillate Oil

Flue gas $= \left(5000 \frac{ft^3}{min}\right)\left(\frac{32 + 460\ °R}{350 + 460\ °R}\right)\left(\frac{lb\text{-}mole}{359\ scf}\right)\left(\frac{60\ min}{hr}\right)$

$= 507.6$ lb-moles/hr

$SO_2$ $= 7.67 \frac{lbs}{hr} \left(\frac{lb\text{-}mole}{64.0\ lb}\right)$

$= .1198$ lb-moles/hr

Fuel S $= 0.5\%$

from Figure 6.3.1-1 % $SO_3$ $= 4.3\%$

The $SO_3$ concentration can be determined by calculation to be

$$[SO_3] \quad = 11 \text{ ppm}$$

One can see in Figure 6.3.1-2 below that the residual oil-fired flue gas will form bisulfate at a higher temperature (earlier in the exchanger). Also, if the $NH_3$ concentration does not become limiting, the greater $SO_3$ concentration will drive the equilibrium of the reaction further to the right, creating more bisulfate and eventually sulfate, $(NH_4)_2SO_4$.

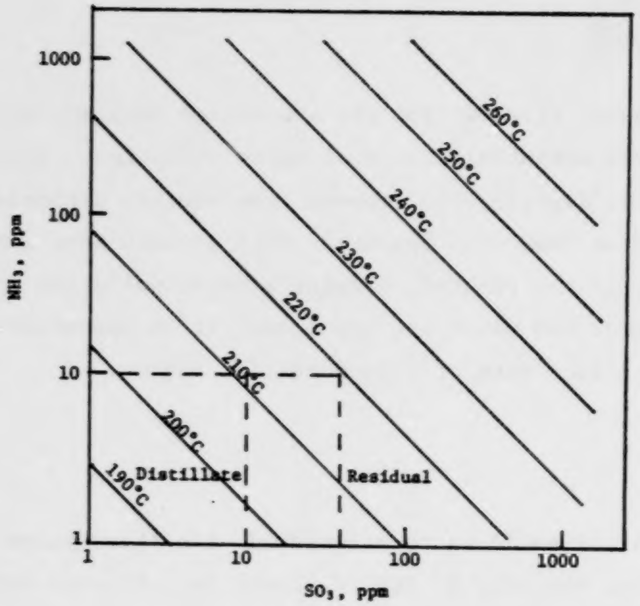


Figure 6.3.1-2. Temperatures below which $NH_4HSO_4$ forms.[26]

Removing the $SO_3$ as bisulfate using particulate control equipment may be a more effective method of removing $SO_3$ from the environment than FGD. This would be a beneficial impact of bisulfate formation. A downstream FGD system could potentially absorb the small $NH_3$ emissions and, therefore, $NH_3$ emissions may be negligible if FGD is used on conjunction with FGT. The level of removal that can be achieved by an FGD scrubber has not yet been examined. Also, the effect of absorbed $NH_3$ on the FGD chemistry has not been resolved, although this question is being studied by the EPA. This is due primarily to the fact that there is only one installation where FGD is applied downstream of an SCR unit and data from this Japanese installation has not been published in the U.S. Several things can be done to reduce the adverse environmental impacts. Combustion control with less $O_2$ minimizes formation of $NO_x$ and $SO_3$. This would be the case for a boiler equipped with low $NO_x$ burners. A minimum NH injection ratio is needed for low $NH_3$ emissions and bisulfate formation. Heat exchanger temperatures must be kept above bisulfate formation and acid condensation points. Use of corrosion-resistant material is warranted where bisulfate deposits are probable. Verticle tube heat exchangers are preferable since they are less prone to plugging.

### 6.3.2 Water Pollution

There are no water streams that are associated with $NO_x$-only SCR systems, however, there is one potential source of water pollution. In some Japanese installations $NH_4HSO_4$ deposits are removed from the air preheater by water washing. The blowdown from this operation will contain both ammonium and sulfate ions which, if not treated, present a water pollution source. Since the amounts of $NH_4HSO_4$ and water are not known, it is impossible to estimate the concentration or flow rate of this potential source.

### 6.3.3 Solid Waste

The only potential solid waste associated with this system is spent catalyst. Presently, the life of SCR catalysts is 1-2 years and the topic of recycling is not addressed in the literature. Since, to date, virtually no commercial SCR units have operated long enough to require catalyst replacement, there is no commercial experience on the treatment of spent catalyst.[27] With the high cost of some of the catalysts, recycling seems to be warranted. Recycling is feasible where the spent catalyst support is composed of valuable materials such as titanium, stainless steel or possibly a ceramic material. Alumina catalyst supports probably do not warrant recycling unless required for environmental reasons. These catalysts can probably be disposed of in the same manner as other industrial catalysts.

The amount of catalyst that must be recycled or disposed of is one reactor volume since replacement involves total catalyst replacement. The actual frequency of catalyst replacement is unique to each specific process, however, catalyst lifetimes are typically one or more years.

Few process vendors have published their catalyst formulations since the field is very competitive at the present time. Base metal oxides are typically used.[24] The environmental impact of catalyst disposal will depend on what materials and compounds are involved.

### 6.3.4 Other Environmental Impacts

The only other environmental impact is a secondary impact. $NH_3$ is commonly made from natural gas and its consumption is considered a secondary environmental impact. This impact will be quantified in a forthcoming report on the impact of $NO_x$ regulations on the $NH_3$ industry. The report is being prepared by TVA under contract to EPA-IERL.

Other adverse environmental impacts (noise, thermal pollution, electrical discharges, etc.) are not present with SCR systems.

### 6.3.5 Environmental Impacts on Modified and Reconstructed Facilities

The environmental impacts of new and retrofitted facilities should be similar. There is not enough difference between new and retrofit systems to indicate that environmental impacts would be significantly different with retrofitted systems.

## 6.4 ENVIRONMENTAL IMPACTS OF CONTROLS FOR GAS-FIRED BOILERS

### 6.4.1 Air Pollution

Emission rates for primary and secondary pollutants are listed in Table 6.4.1-1. The table is organized according to control level (uncontrolled, moderate, and stringent). The impact on the primary pollutant, $NO_x$, is given for each case, as is that of the secondary pollutant, $NH_3$. There is an insignificant amount of particulates in the flue gas and, therefore, these are not considered to be a pollutant. There is also no problem with bisulfate formation since the fuel has only a trace of sulfur.

Table 6.2.1-13 shows the uncontrolled $NO_x$ emission for the natural gas-fired standard boiler to be less than the SIP control level.

TABLE 6.4.1-1. AIR POLLUTION IMPACTS FROM BEST $NO_x$ FGT CONTROL TECHNIQUES FOR GAS-FIRED BOILERS[5,6,7]

| Standard boiler | | | | $NO_x$ | | $NH_3$ | |
|---|---|---|---|---|---|---|---|
| Heat rate (MBtu/hr) | Type | Control level | Type of control | g/s (lb/hr) | ng/J (lb/MBtu) | g/s (lb/hr) | ng/J (lb/MBtu) |
| 15 | Firetube | Uncontrolled | - | 0.332<br>(2.63) | 75.4<br>(.175) | 0 | 0 |
| | | Moderate-70% | SCR Fixed Packed Bed | 0.0995<br>(.789) | 22.6<br>(.0526) | 0.00113<br>(.00898) | 0.257<br>(.000598) |
| | | Stringent-90% | SCR Fixed Packed Bed | .0332<br>(.263) | 7.54<br>(.0175) | 0.00113<br>(.00898) | 0.257<br>(.000598) |
| 150 | Watertube | Uncontrolled | - | 3.31<br>(26.26) | 75.3<br>(.175) | 0 | 0 |
| | | Moderate-70% | SCR Fixed Packed Bed | 0.993<br>(7.88) | 22.6<br>.0525) | 0.00511<br>(.0405) | 0.232<br>(.000540) |
| | | Stringent-90% | SCR Fixed Packed Bed | 0.331<br>(2.63) | 7.53<br>(.0175) | 0.00511<br>(.0405) | 0.232<br>(.000540) |

The only environmental impacts are $NO_x$ and $NH_3$ emissions. The uncontrolled $NO_x$ emissions, 0.332 $\frac{g}{s}$, are the lowest for all standard boilers except distillate oil. Moderate - stringent controls reduce this figure to 0.0995-0.0332 $\frac{g}{s}$. $NH_3$ emissions for all control levels are 1 ppm (Figure 6.1-1). This corresponds to a mass rate of 0.00113 $\frac{g}{s}$.

To reduce the adverse environmental impacts, combustion controls utilizing less $O_2$ minimizes $NO_x$ formation could be implemented. $NH_3$ emissions are presently quite low. Care needs to be taken to see that an excessive $NH_3$:$NO_x$ injection ratio is not used thus increasing the low emission level.

### 6.4.2 Water Pollution

There are no water streams in SCR fixed packed bed systems.

### 6.4.3 Solid Waste

The only potential solid waste associated with this system is spent catalyst. Presently, the life of SCR catalysts is 1-2 years and the topic of recycling is not addressed in the literature. Since, to date, virtually no commercial SCR units have operated long enough to require catalyst replacement, there is no commercial experience on the treatment of spent catalyst.[27] With the high cost of some of the catalysts, recycling seems to be warranted. Recycling is feasible where the spent catalyst support is composed of valuable materials such as titanium, stainless steel, or possibly a ceramic material. Alumina catalyst supports probably do not warrant recycling unless required for environmental reasons. These catalysts can probably be disposed of in the same manner as other industrial catalysts.

The amount of catalyst that must be recycled or disposed of is one reactor volume since replacement involves total catalyst replacement. The actual frequency of catalyst replacement is unique to each specific process, however, catalyst lifetimes are typically one or more years.

Few process vendors have published their catalyst formulations since the field is very competitive at the present time. Base metal oxides are typically used.[24] The environmental impact of catalyst disposal will depend on what materials and compounds are involved.

### 6.4.4 Other Environmental Impacts

The only other environmental impact is a secondary impact. $NH_3$ is commonly made from natural gas and its consumption is considered a secondary environmental impact. This impact will be quantified in a forthcoming report on the impact of $NO_x$ regulations on the $NH_3$ industry. The report is being prepared by TVA under contract to EPA-IERL.

Other adverse environmental impacts (noise, thermal pollution, electrical discharge, etc.) are not present with SCR systems.

### 6.4.5 Environmental Impacts on Modified and Reconstructed Facilities

The environmental impacts of new and retrofitted systems should be similar. Retrofitted systems are not so different as to create a greater adverse environmental impact for these systems.

# REFERENCES

1. Ando, Jumpei. "$NO_x$ Abatement for Stationary Sources in Japan." EPA Report Currently in Preparation, October 1978. p. 3-27.

2. Faucett, H.L., *et al*. Technical Assessment of $NO_x$ Removal Processes for Utility Application. EPA-600/7-77-127. November 1977. p. 240.

3. *Ibid.*, p. 239, 240.

4. Ando, J., *op.cit.*, p. 3-33.

5. *Ibid.*, p. 4-4.

6. Faucett, H.L., *op.cit.*, p. 214.

7. *Ibid.*, p. 259.

8. Ando, J., *op.cit.*, p. 3-7.

9. *Ibid.*, p. 4-41.

10. *Ibid.*, p. 4-95.

11. Wong-Woo, Harmon. "Observation of FGD and Denitrification Systems in Japan." State of California Air Resources Board-SS-78-004. March 7, 1978. Appendix IV. p. 30.

12. *Ibid.*, p. 32.

13. Ando, J., *op.cit.*, p. 4-37.

14. *Ibid.*, p. 4-38.

15. *Ibid.*, p. 4-93.

16. *Ibid.*, p. 4-104.

17. *Ibid.*, p. 4-126.

18. Faucett, H.L., *op.cit.*, p. 5.

19. Ando, J., *op.cit.*, p. 3-47.

20. Ando, J., *op.cit.*, p. 3-55-363.

21. Telephone conversation, Gary Jones of Radian with Jim Turner of EPA-IERL. 10 January 1979.

22. Ando, Jumpei. "$NO_x$ Abatement for Stationary Sources in Japan." EPA-600/7-77-103b. September 1977. p. 47.

23. Ando, J., *op.cit.*, October 1978. pp. 3-61, 3-62.

24. *Ibid.*, p. 3-2.

25. *Ibid.*, p. 3-49.

26. Faucett, H.L., *op.cit.*, p. 302.

27. Ando, J., Letter of April 4, 1979 to J. David Mobley.

# SECTION 7

# EMISSION SOURCE TEST DATA

## 7.1 INTRODUCTION

Test data from operating units are necessary to demonstrate that the control technology will perform as claimed. For this purpose the most meaningful test data are those that represent 24 hour averages over 30 days of continuous operation. At the present time, very little of this type of information is published. However, some continuous data have been presented at recent seminars and obtained from the process operators.

The EPA approved test methods are the same for all fuel types and are discussed here to avoid unnecessary repetition. There are two methods for measuring $NO_x$ (expressed as $NO_2$) in gas streams. First, the EPA Reference Method 7 is for the determination of nitrogen oxides emissions from stationary sources. Presently, this method is the only one approved by the EPA for measuring $NO_x$ levels in flue gas from industrial boilers for emission source test data.

Method 7 is based on grab sampling for wet chemical analysis and is used for spot-checking SCR systems' performance and in calibrating an instrument analyzer. Continuous monitoring by Method 7 for process control purposes is impractical as the method requires a collected sample to set a minimum of 16 hours. However, continuous data can be developed using Method 7 by taking samples at several intervals during a 24 hour period and computing a 24 hour average. Daily values computed in this manner can represent continuous data when computed for a period of 30 days or more.

The second method, EPA Reference Method 20, is for the determination of nitrogen oxide emissions from stationary gas turbines. While this method is not approved for industrial boilers, it is applicable to continuous monitoring due to its utilization of an instrumental analyzer based on chemiluminescence. This instrument provides a sound basis for process control, which is most important when hourly ambient $NO_x$ standards are in effect.

The methodology and test procedures for each method are described in the following paragraphs. With Method 7, grab samples are collected in an evacuated flask containing a dilute sulfuric acid-hydrogen peroxide ($H_2SO_4$-$H_2O_2$) absorbing solution. The nitrogen oxides, except nitrous oxide ($N_2O$), are measured colorimetrically using the phenoldisulfonic acid (PDS) procedure. The apparatus for this system is shown in Figure 7.1-1.

A 25 ml aliquot of absorbing solution is added to the flask. The flask is stoppered and then evacuated by use of the pump. After checking for leaks, the probe and vacuum tube are purged using the squeeze bulb. Then the flask valve is turned to the "sample" position allowing the gas to enter the flask. When the pressures in the flask and sample line (*i.e.*, duct) are equalized the flask is isolated, disconnected from the sampling train, and shaken for 5 minutes. The sample flask is allowed to set for a minimum of 16 hours. After transferring the sample and then washing out the sample flask into a volumetric flask, 25 ml aliquot is pipetted into a porcelain evaporating dish. This aliquot is evaporated to dryness on a steam bath and allowed to cool. Two ml of phenoldisulforic acid solution is added to the dried residue and the residue is ground to a powder with a polyethylene policeman. After adding deionized, distillated water and concentrated sulfuric acid, concentrated ammonium hydroxide is added dropwise until the pH is 10. The contents of the flask are mixed thoroughly and the absorbance of a sample is measured by a spectrophotometer. The total mass of $NO_x$ per sample is expressed by the following equation:

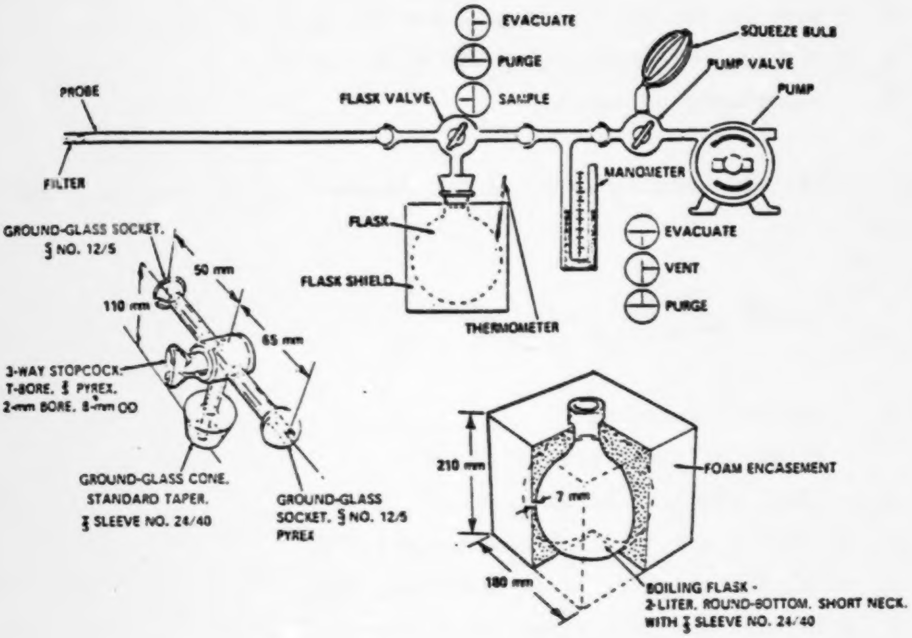

EVACUATE
PURGE
SAMPLE
SQUEEZE BULB
PUMP VALVE
PUMP
PROBE
FLASK VALVE
FILTER
MANOMETER
FLASK
FLASK SHIELD
THERMOMETER
EVACUATE
VENT
PURGE
GROUND-GLASS SOCKET. § NO. 12/5
50 mm
110 mm
65 mm
3-WAY STOPCOCK: T-BORE. § PYREX. 2-mm BORE. 8-mm OD
GROUND-GLASS CONE. STANDARD TAPER. § SLEEVE NO. 24/40
GROUND-GLASS SOCKET. § NO. 12/5 PYREX
210 mm
7 mm
FOAM ENCASEMENT
180 mm
BOILING FLASK - 2-LITER. ROUND-BOTTOM. SHORT NECK. WITH § SLEEVE NO. 24/40

Figure 7.1-1. Sampling train, flask valve, and flask.[1]

$$m = 2\ K_c\ A$$

where $m$ = mass of $NO_x$ as $NO_2$ in gas sample, μg
$K_c$ = spectrophotometer calibration factor
$A$ = absorbance of sample

The sample volume, dry basis, corrected to standard conditions is found by the equation

$$V_s = 0.3858\ \frac{°K}{mmHg}\left(V_f - 25ml\right)\left(\frac{P_f}{T_f} - \frac{P_i}{T_i}\right)$$

where $V_s$ = sample volume, ml
$V_f$ = volume of flask and valve, ml
$P_f$ = final absolute flask pressure, mmHg
$T_f$ = final absolute flask temperature, °K
$P_i$ = initial absolute flask pressure, mmHg
$T_i$ = initial absolute flask temperature, °K

Finally, the $NO_x$ concentration in the gas sample is determined by

$$C = 10^3\ \frac{mg/m^3}{\mu g/ml}\left(\frac{m}{V_s}\right)$$

Method 20, for determining nitrogen oxides emissions from stationary gas turbines, utilizes an instrumental analyzer to which a continuous gas sample from the exhaust stream is conveyed. The apparatus for this system is shown in Figure 7.1-2. Particulate matter and water vapor are the primary interferring species for most instrumental analyzers, but these are removed by the filter and condenser, respectively, present in the sampling train. In application to SCR systems on boilers, the presence of $NH_3$ may interfere with the instruments performance. This problem can be circumvented via the use of an ammonia decomposition catalyst before the probe measuring the reactor outlet $NO_x$ concentration.

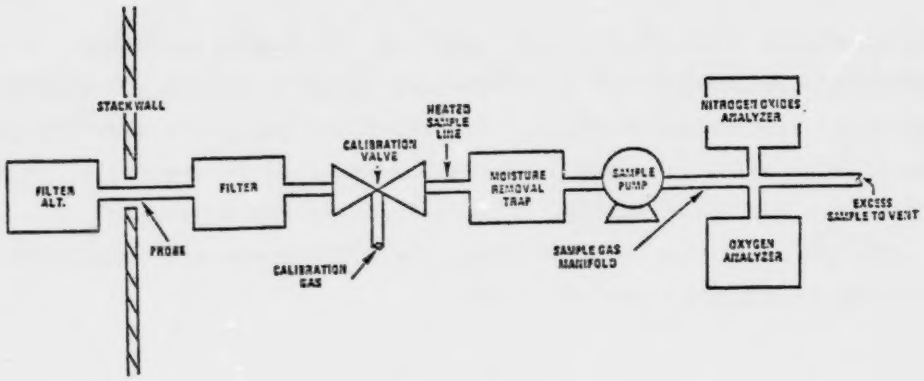

STACK WALL
HEATED SAMPLE LINE
NITROGEN OXIDES ANALYZER
CALIBRATION VALVE
FILTER ALT.
FILTER
MOISTURE REMOVAL TRAP
SAMPLE PUMP
EXCESS SAMPLE TO VENT
PROBE
CALIBRATION GAS
SAMPLE GAS MANIFOLD
OXYGEN ANALYZER

Figure 7.1-2. Measurement system design for stationary gas turbine tests.[2]

## 7.2 EMISSION SOURCE TEST DATA FOR COAL-FIRED BOILERS

At this writing, there have been only a few pilot plant tests performed in Japan utilizing SCR systems to treat coal-fired flue gas. Little data from these tests have been released. Most all SCR work has been done recently in Japan (the oldest SCR system, on an oil-fired boiler, has been operating for 5 years). Because of Japan's lack of large coal reserves, there are few coal-fired utility or industrial boilers in the country. However, more coal-fired boilers are planned for the near future which will utilize imported coal. This is a result of the scarcity and high cost of cleaner fuels.

There are two coal-fired utility boilers equipped with SCR $NO_x$ removal systems due to start-up in 1980.[3] Hokkaido Electric Company plans a 90 MW coal-fired boiler with an SCR unit to be started up at Tomato in March 1980. The Electric Power Development Corporation has a 250 MW coal-fired SCR unit due for completion at Takehara in November, 1980. Also, in the United States, there are 2 SCR pilot plants presently under construction at coal-fired utilities located in Tampa, Florida and Albany, Georgia. These are scheduled for completion in 1979. Once these units are in operation it will be possible to obtain more test data.

The available coal-fired source test data is summarized below. Figure 7.2-1 shows the performance of a cylindrical catalyst treating coal-fired flue gas after particulate removal. Figures 7.2-2 and 7.2-3 show the performance of a parallel flow and a moving bed reactor, respectively, treating coal-fired flue gas after an ESP. Figure 7.2-4 is for a parallel flow system. One can see from the plots that the SCR systems are capable of achieving the stringent level of control.

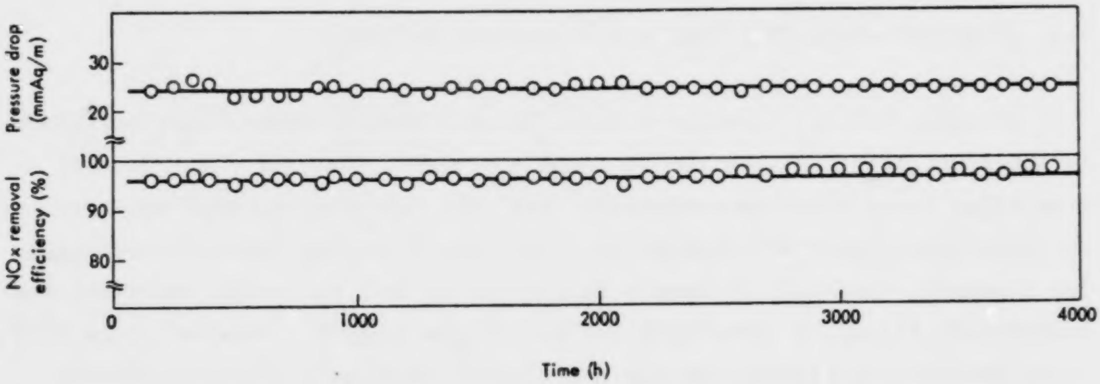


Figure 7.2-1. Change of $NO_x$ removal efficiency and pressure drop (Kawasaki Heavy Industries process, Takehara power station, Hiroshima, Japan).[4]

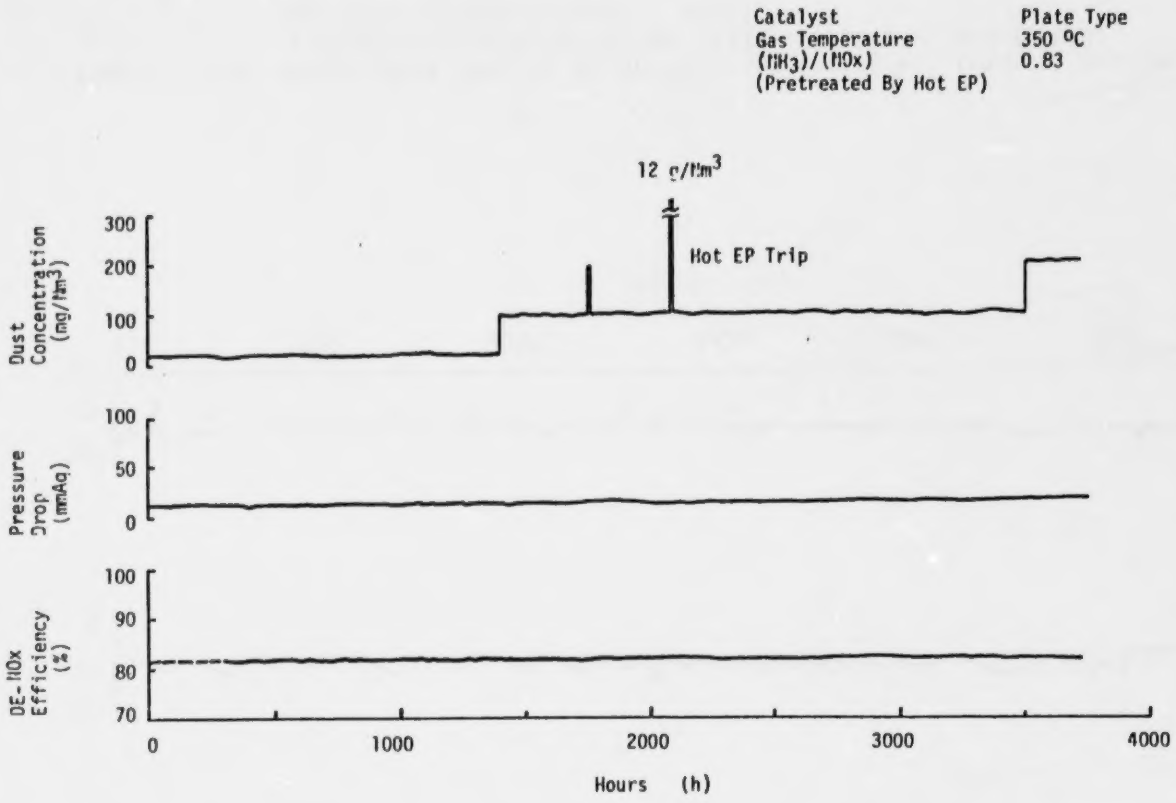


Figure 7.2-2. Pilot plant test of a parallel flow reactor treating a flue gas from a coal-fired utility boiler (Hitachi, Ltd. process, unknown location, Japan).[5]

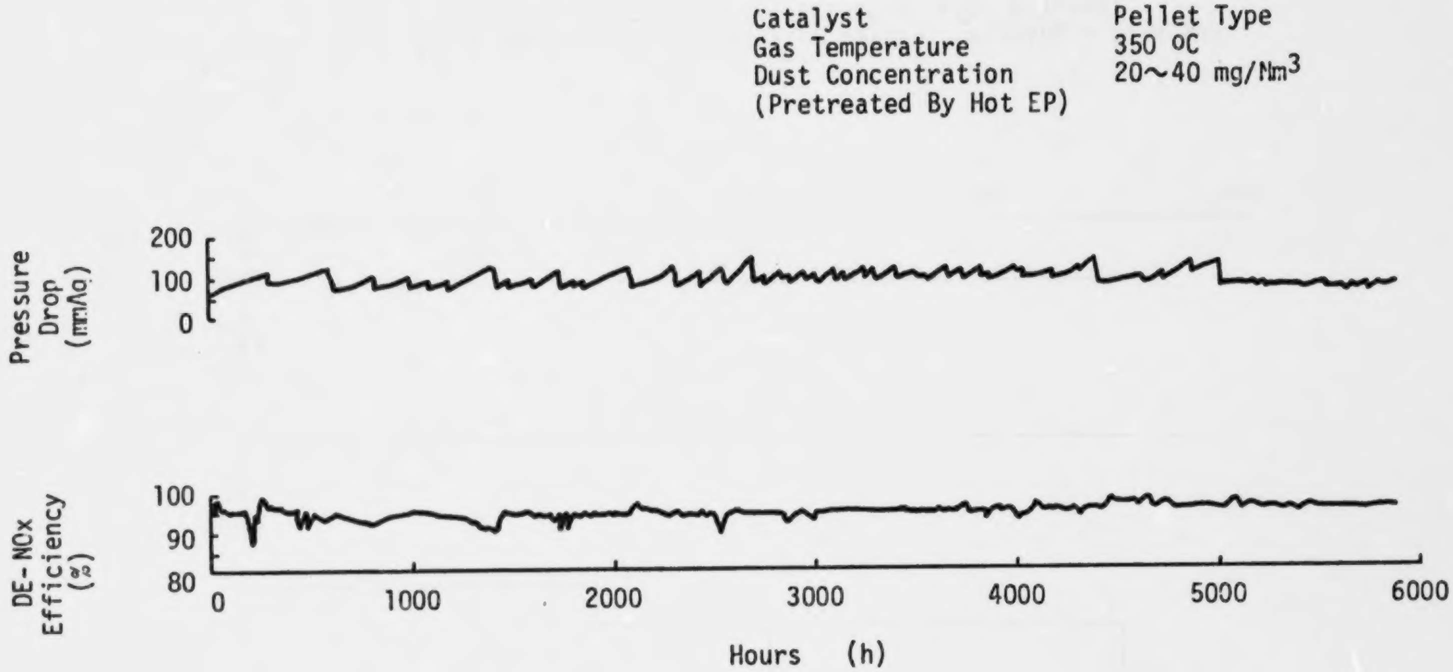


Figure 7.2-3. Pilot plant test of an intermittent-moving bed reactor treating a flue gas from a coal-fired utility boiler (Hitachi, Ltd. process, unknown location, Japan).[6]

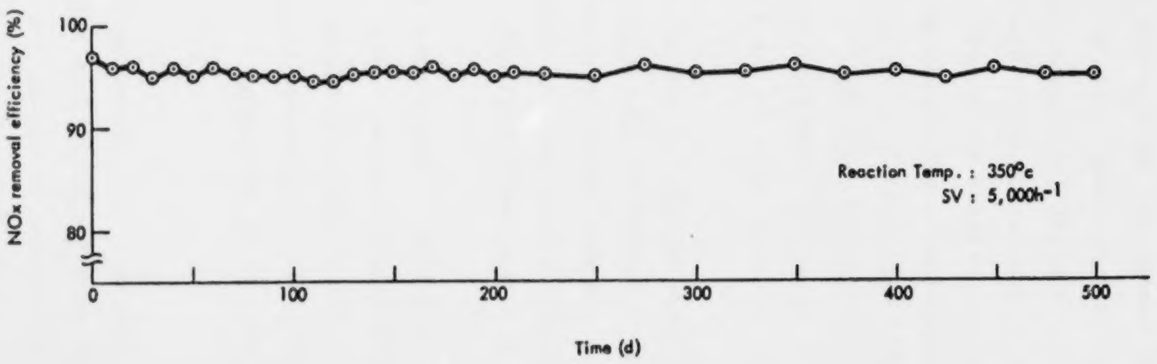


Figure 7.2-4. Durability test of $NO_x$ removal catalyst (Kawasaki Heavy Ind. process, Takehara power station, Hiroshima, Japan).

## 7.3 EMISSION SOURCE TEST DATA FOR OIL-FIRED BOILERS

While there are a number of commercial SCR systems presently treating oil-fired flue gas in Japan, the data on these units are limited mostly to a single reported removal level. Catalyst life tests on heavy oil-fired pilot unit equipment do provide an indication as to how commercial units will perform. These continuous test results are shown in Figures 7.3-1 through 7.3-3. Most of the data available are presented as summaries of pilot test results and are usually expressed in plots of $NO_x$ removal (%) *vs.* $NH_3$:$NO_x$ mole ratio or reactor space velocity ($hr^{-1}$). In a few cases, tables of operating parameters of commercial SCR plants are given. These results are given in Tables 7.3-1 through 7.3-3 on the following pages.

Most data give only point values of removal and not a set of continuous data. In addition, the test method and boiler operating conditions are not given. Included in these figures are data recently obtained from commercial Japanese installations on industrial boilers and, as such, they represent the most complete set of continuous data currently available.

Summaries of the oil-fired industrial (larger than 3 MW) and utility SCR plants in Japan are shown in Tables 7.3-4 and 7.3-5. These tables are presented since they represent locations where operating data on SCR units can be obtained. The data can be obtained by either contacting the boiler owners and requesting available data or by arranging for independent on-site testing.

The data shown in Figures 7.3-4 and 7.3-5 represent the most recently available continuous daily average data available from SCR systems applied to industrial boilers. EPA Method 20 was used to obtain the data in both of the figures. The average removal level represents the level of control necessary to meet the local Japanese emission regulations. Continuous daily averages are not available from most SCR units; however, Table 7.3-6 shows maximum and minimum $NO_x$ removal values for several industrial units.

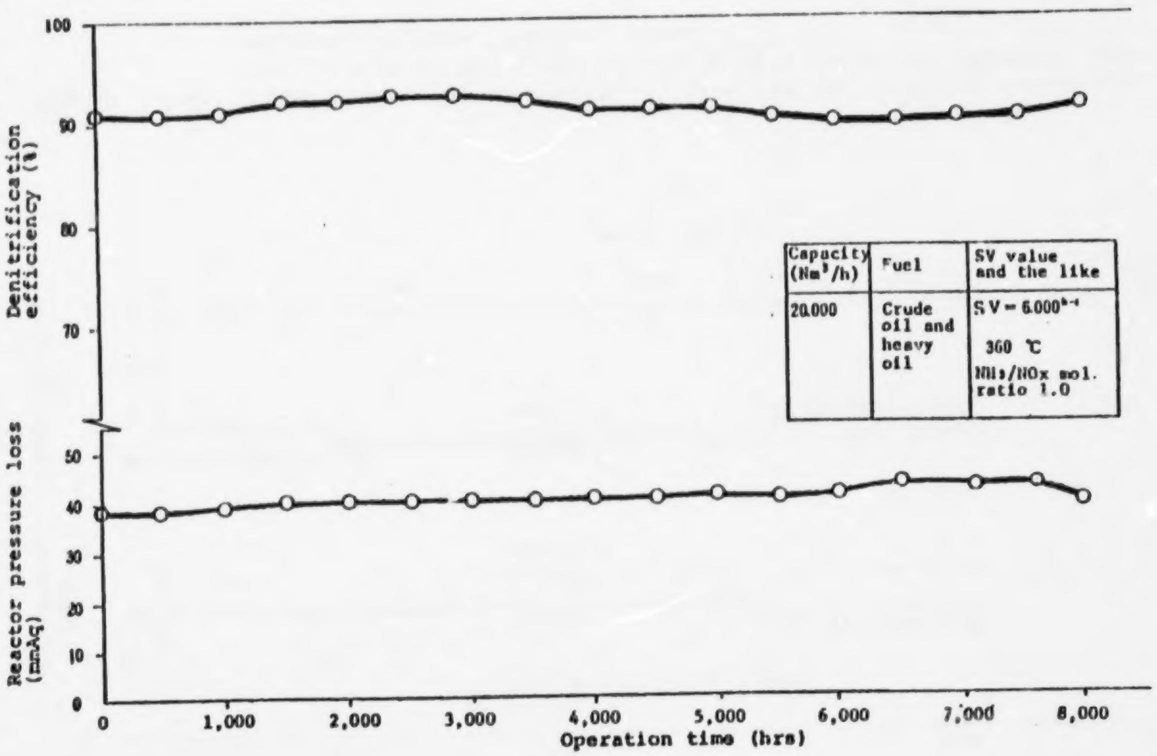


Figure 7.3-1. Catalyst life test results (IHI process, Taketoyo power station, Japan).[7]

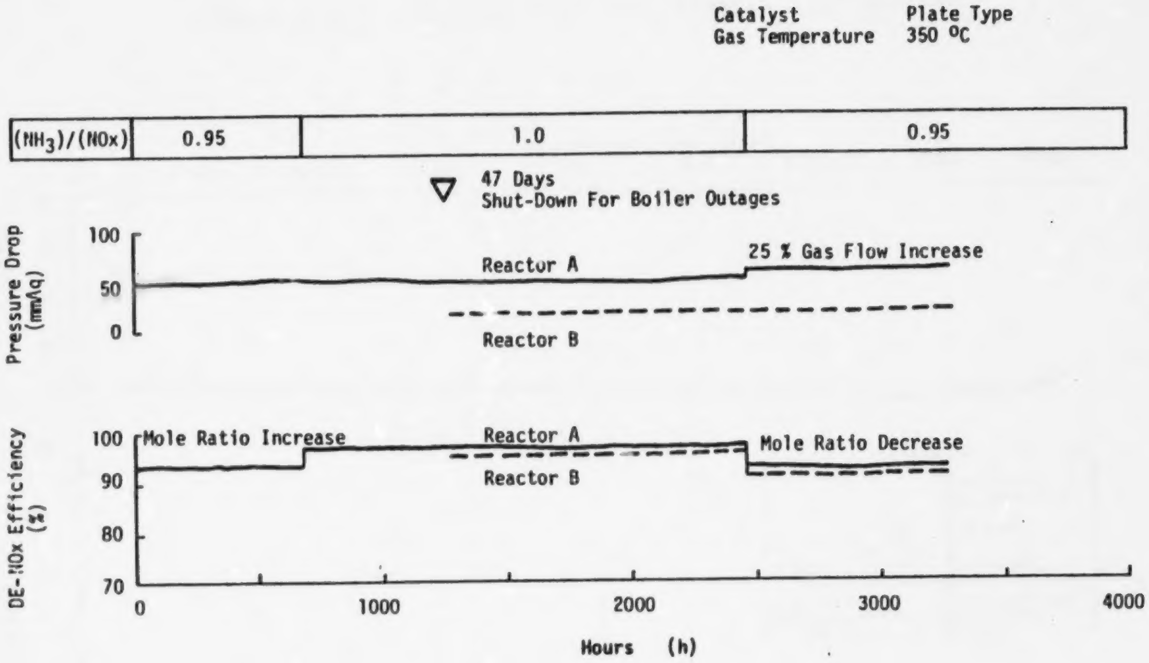
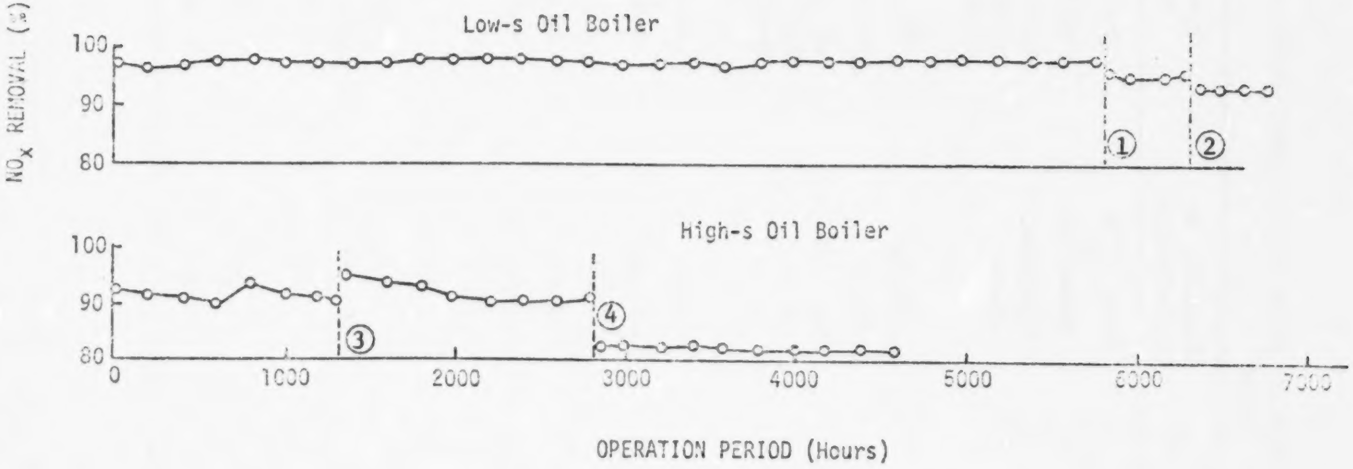


Figure 7.3-2. Pilot plant test of a parallel flow reactor treating a flue gas from a high sulfur heavy oil-fired utility boiler (Hitachi, Ltd. process, unknown location, Japan).[9]



Circled figures show times when SV and $NH_3$/NO mole ratio were changed.

1. SV 10,000 - 15,000 $hr^{-1}$
2. SV 15,000 - 20,000 $hr^{-1}$
3. SV 6,200 - 4,500 $hr^{-1}$
4. SV 4,500 - 6,200 and the mole ratio 0.95 - 0.83.

Figure 7.3-3. Test results of oil-fired boilers (Hitachi, Ltd. process, unknown location, Japan).[16]

TABLE 7.3-1. OPERATION PARAMETERS OF MAJOR PLANTS CONSTRUCTED BY HITACHI ZOSEN[10]

| | Idemitsu Kosan | Shindaikyowa Petrochemical | Kawasaki Steel |
|---|---|---|---|
| Completed | Oct. 1975 | Nov. 1975 | Nov. 1976 |
| Plant site | Chiba | Yokkaichi | Chiba |
| Gas source | FCC-CO Boiler and furnace | Oil-fired Boiler | Iron-ore Sintering machine |
| Capacity ($Nm^3/hr$) | 350,000 | 440,000 | 762,000 |
| Load factor (%) | 50-100 | 50-100 | 70-100 |
| Pretreatment of gas | Heating | EP*, FGD, Heating | EP, FGD WEP†, Heating |
| Reactor inlet | | | |
| $NO_x$ (ppm) | 230 | 150 | 200-300 |
| $SO_x$ (ppm) | 50-80 | 80-130 | 5-20 |
| Dust ($mg/Nm^3$) | 20-50 | 30-100 | 3-10 |
| $O_2$ (%) | 2.3 | 3.2 | 11.2 |
| Reactor type | Fixed bed | Fixed bed | Fixed bed |
| Reaction temp. | 400 | 420 | |
| $NO_x/NH_3$ ratio | 1.0 | 1.0 | 1.0 |
| Catalyst No. | 204 | 304 | 304 |
| SV ($hr^{-1}$) | 5,000 | 10,000 | 4,000 |
| $NO_x$ removal (%) | 93 | 80‡ | 95 |
| Pressure drop by SCR reactor ($mmH_2O$) | 170 | 160 | 50 |
| Catalyst life | 1 year | 1 year | 1 year |

*Electrostatic precipitator

†Wet electrostatic precipitator

‡Including leakage in heat exchanger

TABLE 7.3-2. SCR PLANTS BY MITSUI ENGINEERING & SHIPBUILDING CO.[11]

| | Mitsui Petro-Chemical Co. | Ukishima Petro-Chemical Co. |
|---|---|---|
| Capacity ($Nm^3/hr$) | 200,000 | 220,000 |
| Gas composition | | |
| $NO_x$ (ppm) | 190 | 150 |
| $SO_x$ (ppm) | None | 300 |
| Dust ($mg/Nm^3$) | 20-50 | 100-150 |
| Catalyst and reactor | | |
| Catalyst carrier | $Al_2O_3$ | $TiO_2$ |
| Catalyst shape | Granule | Tube |
| SV ($hr^{-1}$) | 2,600 | 4,000 |
| Temperature (°C) | 350 | 350-400 |
| $NH_3/NO_x$ mole ratio | 1.0 | 1.0 |
| $NO_x$ removal (%) | Above 90 | Above 90 |
| Total pressure drop ($mmH_2O$) | | 180 |
| Leak $NH_3$ (ppm) | | Below 10 |
| Operation start | Oct. 1975 | July 1977 |
| Plant cost ($10^6$ yen) | | 260 |
| Denitrification cost (yen/kWhr)* | | |

*Including 7 years depreciation.

TABLE 7.3-3. OPERATION DATA OF SCR PLANTS FOR DIRTY GAS[12]

| | | Commercial | |
|---|---|---|---|
| | Pilot | No. 1 | No. 2 |
| Gas for SCR ($NM^3$/hr) | 30,000 | 240,000 | 300,000 |
| Fuel | Oil(S=0.7%) | Oil(S=0.7%) | (Oil(S=0.7%) |
| Load fluctuation | 60-100% | 60-100% | 60-100% |
| Stack height (m) | 70 | 140 | 140 |
| Inlet gas composition | | | |
| $O_2$ (%) | 6 | 6 | 6 |
| $SO_x$ (ppm) | 400 | 400 | 400 |
| $NO_x$ (ppm) | 200 | 200 | 200 |
| Particulates after EP ($mg/Nm^3$) | 5-20 | 5-10 | 10-20 |
| FGD unit | None | Scheduled | Scheduled |
| SV ($hr^{-1}$) | 5,000 | 5,000 | 5,000 |
| Temperature (°C) | 320 | 320 | 320 |
| $NO_x$ removal (%) | Over 90 | Over 90 | Over 90 |
| $NH_3$/NO mole ratio | 1.0 | 1.0 | 1.0 |
| Leak ammonia (ppm) | 10-20 | 10-20 | 10-20 |
| Type of reactor | Fixed bed | Fixed bed | Moving bed |
| Pressure drop ($mmH_2O$) | | | |
| Reactor | | 200 | |
| Total system | | 500 | |
| Plant completed | July 1973 | Mar. 1976 | Oct. 1976 |

TABLE 7.3-4. OIL-FIRED INDUSTRIAL SCR PLANTS[13]

| Company | Site | Capacity ($Nm^3/hr$) | Reactor type* | Completion date |
|---|---|---|---|---|
| Sumitano Chemical | Sodegaura | 30,000 | FPB | July 1973 |
| Kurabo | Hirakata | 30,000 | MB | August 1975 |
| Mitsui Petrochemical | Chiba | 200,000 | FPB | October 1975 |
| Idemitsu Kosan | Chiba | 350,000 | FPB | October 1975 |
| Shindaikyowa P.C. | Yokkaichi | 440,000 | FPB | November 1975 |
| Sumitomo Chemical | Sodegaura | 240,000 | FPB | March 1976 |
| Fuji Oil | Chiba | 70,000 | PF | July 1976 |
| Sumitomo Chemical | Sodegaura | 300,000 | MB | September 1976 |
| Sumitomo Chemical | Sodegaura | 300,000 | MB | October 1976 |
| Nisshin Steel | Amagasaki | 19,000 | FPB | July 1977 |
| Nisshin Steel | Amagasaki | 20,000 | FPB | August 1977 |
| Chiyoda Kenzai | Kaizuka | 15,000 | MB | October 1977 |
| Fuji Oil | Sodegaura | 200,000 | PF | January 1978 |
| Ajinomoto | Kawasaki | 180,000 | PF | January 1978 |
| Nippon Oils & Fats | Amagasaki | 20,000 | MB | April 1978 |
| Nippon Yakin | Kawasaki | 14,000 | FPB | July 1978 |

*FPB = Fixed Packed Bed
MB = Moving Bed
PF = Parallel Flow

TABLE 7.3-5. OIL-FIRED UTILITY SCR PLANTS[14]

| Power company | Site | Capacity ($Nm^3/hr$) | Reactor type* | Completion date |
|---|---|---|---|---|
| Kansai Electric | Kainan | 300,000 | FPB | June 1977 |
| Company C | - | 1,010,000 | PF | February 1978 |
| Kansai Electric | Amagasaki | 410,000 | - | June 1978 |
| Company A | - | 490,000 | PF | June 1978 |
| Kansai Electric | Osaka | | PF | July 1978 |
| Company D | - | 490,000 | PF | July 1978 |
| Company G | - | 1,000,000 | PF | April 1979 |
| Chugoku Electric | Kudamatsu | 1,900,000 | PF | July 1979 |
| Chubu Electric | Chita | 1,920,000 | PF | February 1980 |
| Tohoku Electric | Niigata | 1,600,000 | PF | August 1981 |

*FPB = Fixed Packed Bed
PF = Parallel Flow

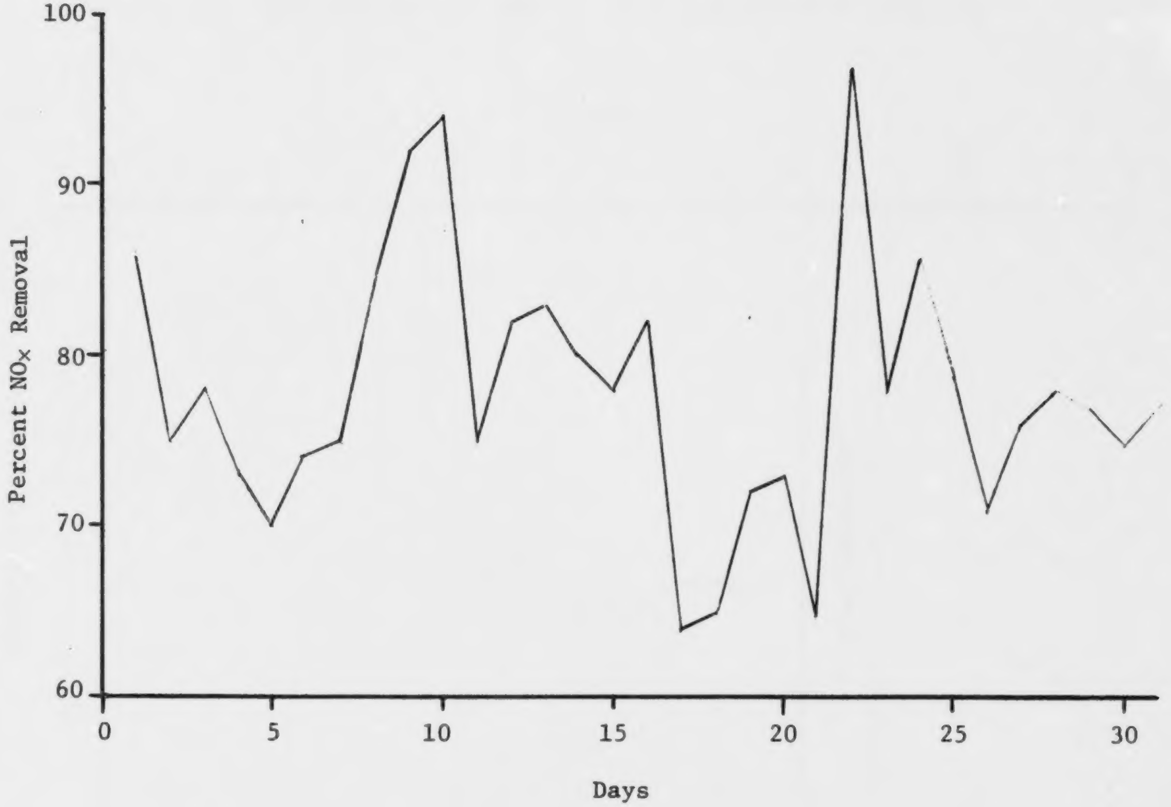


Figure 7.3-4. $NO_x$ removal for the month of May 1977 (Hitachi Zosen fixed bed process, Shindaikyowa Petrochemical, Yokkaichi, Japan, Chemiluminescence Method).[20]

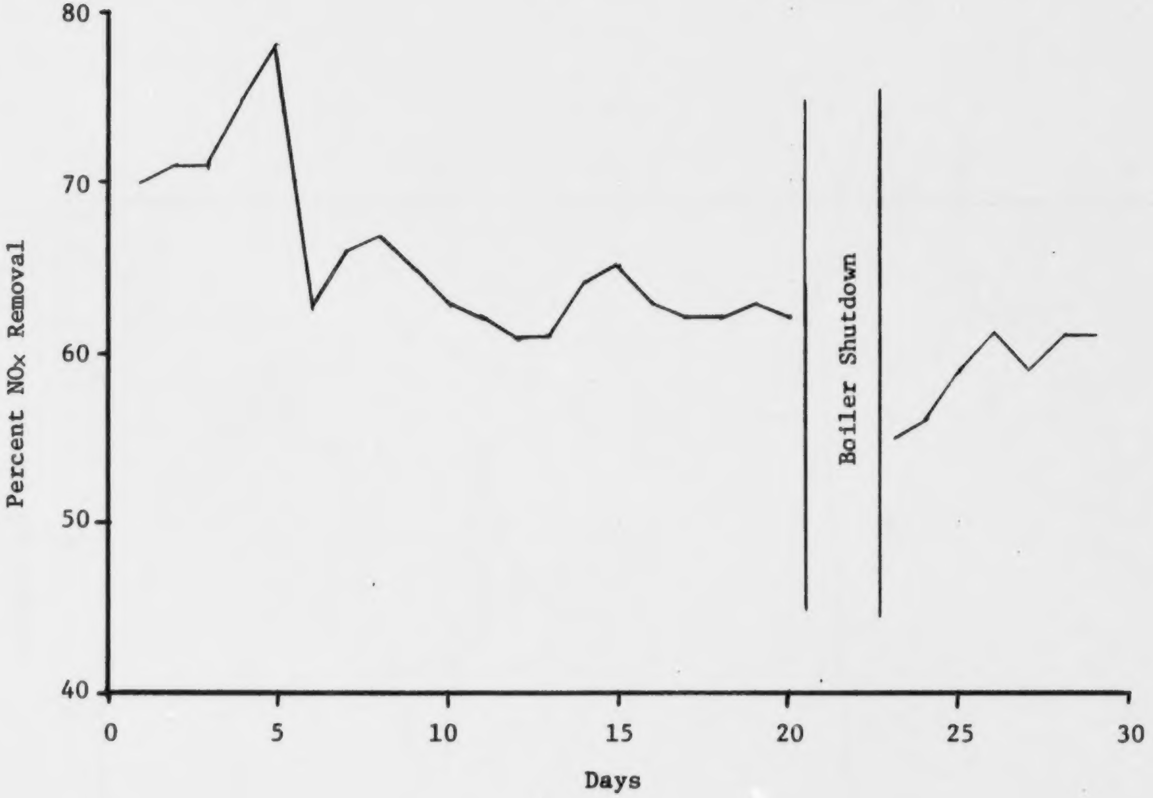


Figure 7.3-5. $NO_x$ removal for August, 1978 (MHI process, Fuji Oil, Sodegaura, Japan, PDS/chemiluminescence method).[20]

TABLE 7.3-6. $NO_x$ REMOVAL LEVELS AT SEVERAL JAPANESE INDUSTRIAL BOILERS WITH $NO_x$ CONTROL BY SCR[20]

| Plant owner | Plant site | Capacity* (MW) | Process | Percent NO removal range |
|---|---|---|---|---|
| Nippon Yakin | Kawasaki | 5 | Fixed bed + sodium scrubbing | 86-98 |
| Shindaikyowa | Yokkaichi | 135 | Sodium scrubbing + fixed bed | 53-62 |
| Fuji Oil | Sodegaura | 23 | Fixed (parallel passage) bed | 93-97 |
| Fuji Oil | Sodegaura | 53 | Fixed (honeycomb) bed | 60-80 |
| Kurabo | Hirakata | 10 | Moving bed (continuous) | 90-94 |
| Nippon Oil & Fats | Amagasaki | 7 | Moving bed (intermittent) | 94-97 |
| Kansai Paint | Amagasaki | 5.3 | Fixed bed | 90-92 |
| Nisshin Steel | Amagasaki | 6.5 | Moving bed | 94-96 |

*Assumed to be $MW_e$ equivalent.

## 7.4 EMISSION SOURCE TEST DATA FOR GAS-FIRED BOILERS

Although gas-fired boilers, both industrial and utility, are numerous in Japan, few have been equipped with SCR units so far. This is due to the fact that combustion modifications on the boilers have been installed because of their lower cost and the lack of fuel-bound nitrogen to contend with. The data available on gas-fired SCR systems in Japan are presented in Figures 7.4-1 through 7.4-5 on the following pages. Figure 7.4-1 is a plot of a long-term performance test while the rest are summaries of pilot tests.

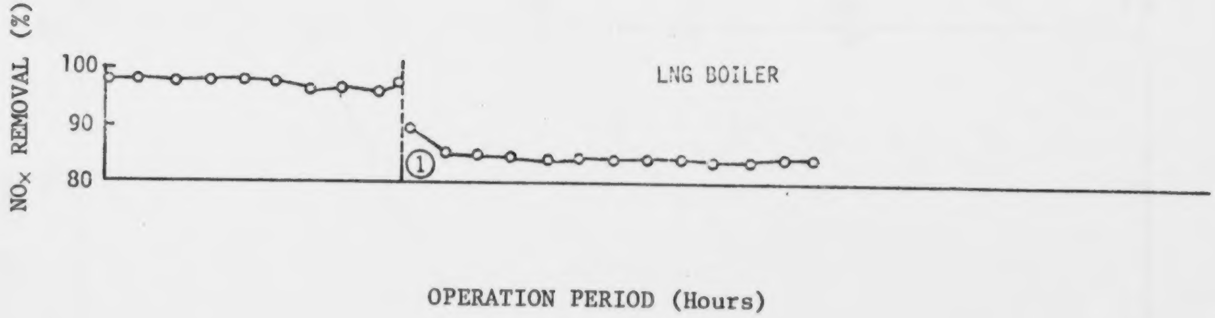
Summaries of the gas-fired industrial and utility SCR plants in Japan are shown in the table below.

TABLE 7.4-1. GAS-FIRED SCR PLANTS[15]

| Company | Site | Capacity ($Nm^3$/hr) | Reactor type* | Completion date |
|---|---|---|---|---|
| Osaka Gas | Takaishi | 15,000x2 | FPB | December 1976 |
| Chubu Electric | Chita | 1,910,000 | FPB | April 1978 |
| Kyushu Electric | Kokura | 1,610,000 | FPB | July 1978 |
| Chubu Electric | Chita | 1,910,000 | FPB | September 1978 |
| Hyushu Electric | Kokura | 1,610,000 | FPB | December 1978 |

*FPB = Fixed Packed Bed

As with oil-fired installations, the type of test data desired may exist, but has not yet been published. To obtain this data it will be necessary to contact the boiler owners to possibly get available data or conduct on-site testing.



Circled figure shows time when SV and $NH_3$/NO mole ratio were changed.

1. SV 10,000 - 20,000 $hr^{-1}$

Figure 7.4-1. Test results of gas-fired boilers (Hitachi, Ltd. process, unknown location, Japan).[16]

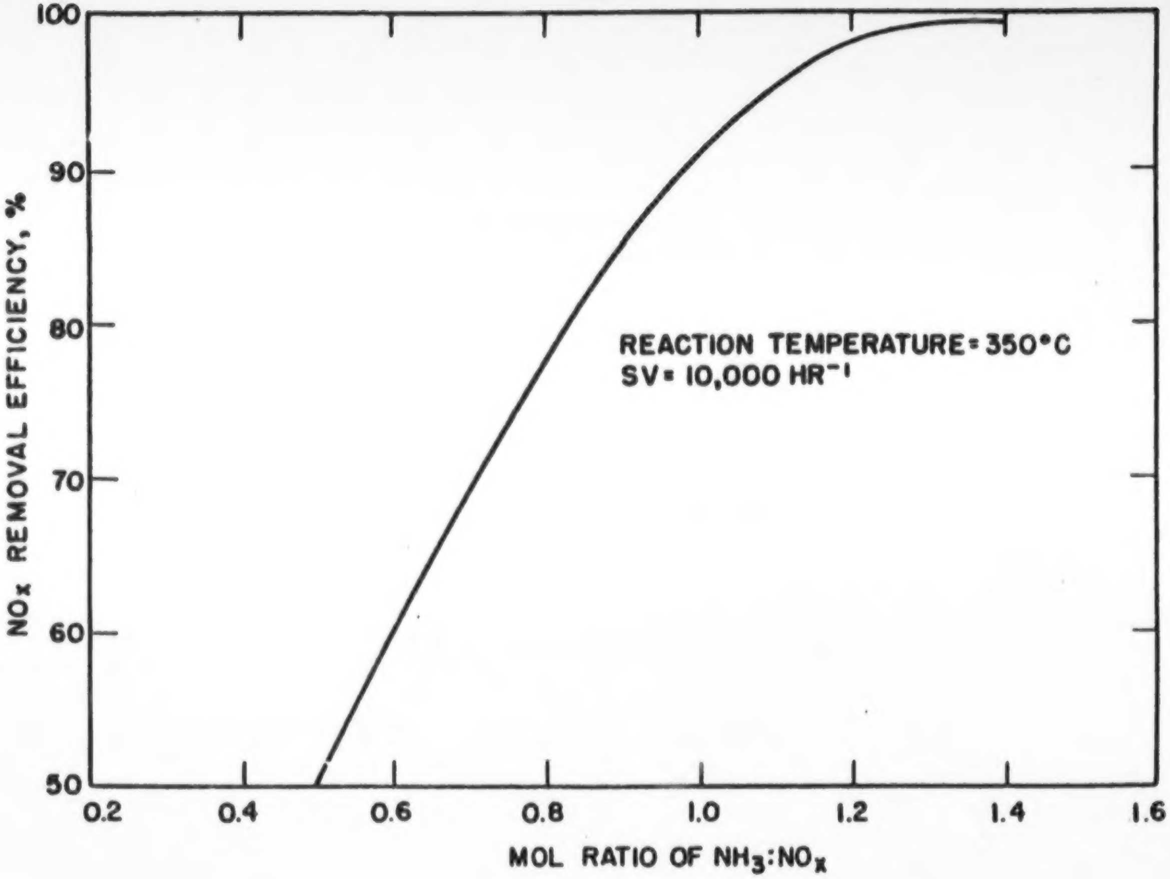


Figure 7.4-2. Characteristic curve of the effect of mole ratio of $NH_3$:$NO_x$ on $NO_x$ removal efficiency for Hitachi, Ltd. process.[17]

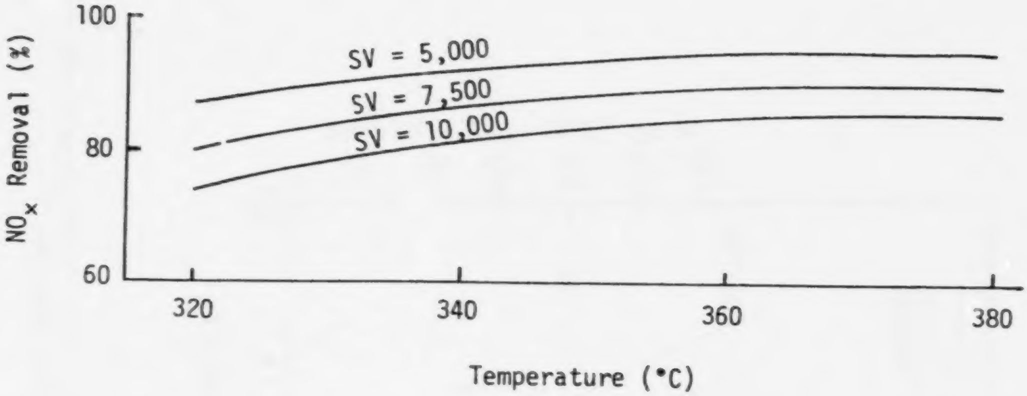


Figure 7.4-3. Performance of catalyst MTC-102 (Mitsui Toatsu process, unknown location, Japan).[18]

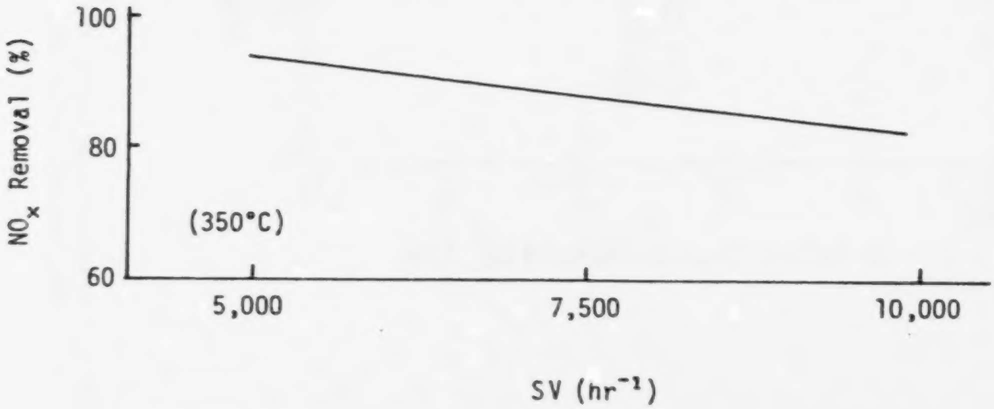


Figure 7.4-4. SV and $NO_x$ removal (MTC-102) (Mitsui Toatsu process, unknown location, Japan).[18]

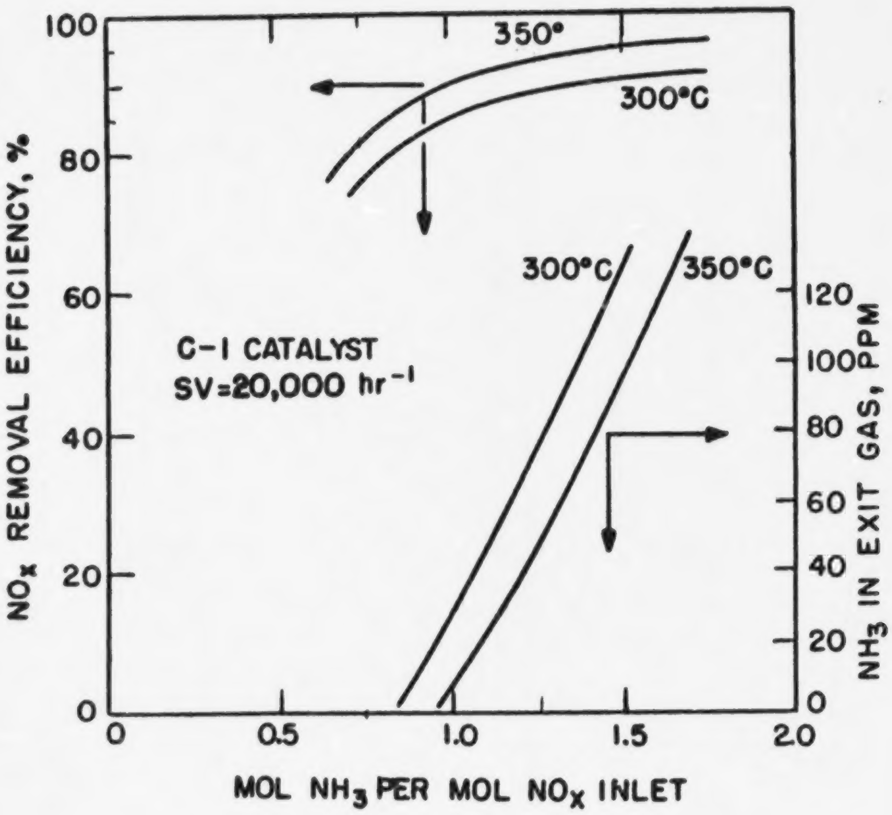


Figure 7.4-5. Relationship among inlet $NH_3$:$NO_x$ mol ratio, $NO_x$ removal efficiency, and exiting $NH_3$ concentration using the Sumitomo Chemical C-1 Catalyst.[19]

## REFERENCES

1. Environmental Reporter, Appendix A, Oct. 21, 1977, p. 92.

2. Federal Register, Volume 42, Number 191, Oct. 3, 1977, p. 53790.

3. Ando, J. $NO_x$ Abatement from Stationary Sources in Japan. EPA draft report in preparation. October 1978, pp. 1-35.

4. Niwa, Senji. Characteristic of Cylindrical $DeNO_x$ Catalyst for a Coal-Fired Boiler. Paper presented at EPRI $NO_x$ Control Technology Seminar II. Denver, Colorado. November 8-9, 1978. p. 21-8.

5. Kuroda, H., Nakajima, F. Experiences of $NO_x$ Removal in Pilot Plants and Utility Boilers. Paper presented at EPRI $NO_x$ Control Technology Seminar II. Denver, Colorado. November 8-9, 1978. p. 20-13.

6. *Ibid.*, p. 20-10.

7. Ando, J., *op. cit.*, p. 4-96.

8. Niwa, S., *op. cit.*, p. 21-5.

9. Kuroda, H., *et al.*, *op. cit.*, p. 20-12.

10. Ando, J., *op. cit.*, p. 4-21.

11. *Ibid.*, p. 4-71.

12. *Ibid.*, p. 4-5.

13. *Ibid.*, p. 3-4, 3-5.

14. *Ibid.*, p. 1-35

15. *Ibid.*, p. 3-4, 3-5.

16. *Ibid.*, p. 4-43.

17. Faucett, H. L., *et al.* Technical Assessment of $NO_x$ Removal Processes for Utility Applications. EPA 600/7-77-127. November 1977. p. 214.

18. Ando, J., *op cit.*, p. 4-121.

19. Faucett, H. L., *op. cit.*, p. 298.

# APPENDIX 1

## DETAILED SYSTEM EVALUATIONS

TABLE A1.1. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: COAL-FIRED BOILERS - MODERATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 8 | 6 | 7 | 11 | 7 | 3 | 9 | 10 | 3 | 5 | 69 |
| SCR Moving Bed | 8 | 4 | 7 | 14 | 7 | 3 | 9 | 10 | 3 | 5 | 70 |
| SCR Parallel Flow | 12 | 7 | 7 | 14 | 8 | 3 | 14 | 10 | 3 | 5 | 83 |
| Absorption - Oxidation | 7 | 2 | 6 | 2 | 3 | 2 | 7 | 6 | 3 | 5 | 43 |

TABLE A1.2. POINT VALUE SIMULTANEOUS $NO_x/SO_x$ PROCESS RATINGS: COAL-FIRED BOILERS - MODERATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Parallel Flow | 12 | 3 | 9 | 11 | 5 | 3 | 12 | 13 | 2 | 2 | 72 |
| Adsorption | * | | | | | | | | | | |
| Electron Beam Radiation | 4 | 5 | 3 | 6 | 4 | 2 | 6 | 3 | 3 | 5 | 41 |
| Absorption - Reduction | 10 | 2 | 7 | 10 | 4 | 4 | 7 | 3 | 3 | 2 | 52 |
| Oxidation - Absorption/Reduction | 10 | 3 | 6 | 6 | 1 | 2 | 8 | 10 | 3 | 2 | 51 |
| Oxidation - Absorption | 8 | 4 | 6 | 5 | 1 | 2 | 8 | 11 | 3 | 5 | 51 |

*Not Applicable - Does not meet removal requirements.

TABLE A1.3. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: COAL-FIRED BOILERS - STRINGENT CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 4 | 6 | 7 | 8 | 7 | 3 | 9 | 10 | 3 | 5 | 62 |
| SCR Moving Bed | 4 | 4 | 7 | 8 | 8 | 3 | 9 | 10 | 3 | 5 | 60 |
| SCR Parallel Flow | 8 | 7 | 7 | 8 | 8 | 3 | 14 | 10 | 3 | 5 | 73 |
| Absorption - Oxidation | * | | | | | | | | | | |

*Not Applicable - Does not meet removal requirements

TABLE A1.4. POINT VALUE SIMULTANEOUS $NO_x/SO_x$ PROCESS RATINGS:
COAL-FIRED BOILERS - STRINGENT CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Parallel Flow | 8 | 3 | 9 | 11 | 5 | 3 | 12 | 13 | 2 | 2 | 68 |
| Adsorption | * | | | | | | | | | | |
| Electron Beam Radiation | * | | | | | | | | | | |
| Absorption - Reduction | * | | | | | | | | | | |
| Oxidation - Absorption/Reduction | 6 | 3 | 6 | 6 | 1 | 2 | 8 | 10 | 3 | 2 | 48 |
| Oxidation - Absorption | 6 | 4 | 6 | 5 | 1 | 2 | 8 | 11 | 3 | 5 | 49 |

*Not Applicable - Does not meet removal requirements.

## TABLE A1.5. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: COAL-FIRED BOILERS - INTERMEDIATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 6 | 6 | 7 | 11 | 7 | 3 | 9 | 10 | 3 | 5 | 67 |
| SCR Moving Bed | 6 | 4 | 7 | 14 | 8 | 3 | 9 | 10 | 3 | 5 | 69 |
| SCR Parallel Flow | 10 | 7 | 7 | 14 | 8 | 3 | 14 | 10 | 3 | 5 | 81 |
| Absorption - Oxidation | 7 | 2 | 6 | 2 | 3 | 2 | 7 | 6 | 3 | 5 | 43 |

## TABLE A1.6. POINT VALUE SIMULTANEOUS $NO_x/SO_x$ PROCESS RATINGS: COAL-FIRED BOILERS - INTERMEDIATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Parallel Flow | 10 | 3 | 9 | 11 | 5 | 3 | 12 | 13 | 2 | 2 | 70 |
| Adsorption | * | | | | | | | | | | |
| Electron Beam Radiation | 4 | 5 | 3 | 6 | 4 | 2 | 6 | 3 | 3 | 5 | 41 |
| Absorption - Reduction | 8 | 2 | 7 | 10 | 4 | 4 | 7 | 3 | 3 | 2 | 50 |
| Oxidation - Absorption/Reduction | 8 | 3 | 6 | 6 | 1 | 2 | 8 | 10 | 3 | 2 | 49 |
| Oxidation - Absorption | 6 | 4 | 6 | 5 | 1 | 2 | 8 | 11 | 3 | 5 | 46 |

*Not Applicable - Does not meet removal requirements.

## TABLE A1.7. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: OIL-FIRED BOILERS - MODERATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 8 | 6 | 7 | 14 | 8 | 3 | 11 | 16 | 3 | 5 | 81 |
| SCR Moving Bed | 12 | 5 | 7 | 14 | 9 | 3 | 14 | 16 | 3 | 5 | 88 |
| SCR Parallel Flow | 12 | 7 | 7 | 14 | 9 | 3 | 14 | 16 | 3 | 5 | 90 |
| Absorption - Oxidation | 7 | 3 | 6 | 2 | 4 | 2 | 8 | 13 | 3 | 5 | 53 |

## TABLE A1.8. POINT VALUE SIMULTANEOUS $NO_x/SO_x$ PROCESS RATINGS: OIL-FIRED BOILERS - MODERATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Parallel Flow | 12 | 4 | 9 | 11 | 7 | 3 | 12 | 13 | 2 | 2 | 75 |
| Adsorption | * | | | | | | | | | | |
| Electron Beam Radiation | 4 | 5 | 3 | 6 | 4 | 2 | 8 | 7 | 3 | 5 | 47 |
| Absorption - Reduction | 8 | 3 | 7 | 9 | 5 | 4 | 7 | 10 | 3 | 2 | 58 |
| Oxidation - Absorption/Reduction | 10 | 4 | 6 | 6 | 2 | 2 | 8 | 16 | 3 | 2 | 59 |
| Oxidation - Absorption | 8 | 5 | 6 | 5 | 2 | 2 | 9 | 7 | 3 | 5 | 52 |

*Not Applicable - Does not meet removal requirements.

## TABLE A1.9. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: OIL-FIRED BOILERS - STRINGENT CONTROL

| | Performance | Operation and maintenance | Environ-mental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Relia-bility | Status of development | Adapt-ability | Compati-bility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 4 | 6 | 7 | 8 | 8 | 3 | 11 | 16 | 3 | 5 | 74 |
| SCR Moving Bed | 8 | 5 | 7 | 8 | 9 | 3 | 14 | 16 | 3 | 5 | 81 |
| SCR Parallel Flow | 8 | 7 | 7 | 8 | 9 | 3 | 14 | 16 | 3 | 5 | 83 |
| Absorption - Oxidation | * | | | | | | | | | | |

*Not Applicable - Does not meet removal requirements

## TABLE A1.10. POINT VALUE SIMULTANEOUS $NO_x/SO_x$ PROCESS RATINGS: OIL-FIRED BOILERS - STRINGENT CONTROL

| | Performance | Operation and maintenance | Environ-mental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Relia-bility | Status of development | Adapt-ability | Compati-bility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Parallel Flow | 8 | 4 | 9 | 11 | 7 | 3 | 12 | 13 | 2 | 2 | 71 |
| Adsorption | * | | | | | | | | | | |
| Electron Beam Radiation | * | | | | | | | | | | |
| Absorption - Reduction | * | | | | | | | | | | |
| Oxidation - Absorption/Reduction | 6 | 4 | 6 | 6 | 2 | 2 | 8 | 16 | 3 | 2 | 54 |
| Oxidation - Absorption | 6 | 5 | 6 | 5 | 2 | 2 | 9 | 7 | 3 | 5 | 50 |

*Not Applicable - Does not meet removal requirements

TABLE A1.11. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: OIL-FIRED BOILERS - INTERMEDIATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 6 | 6 | 7 | 14 | 8 | 3 | 11 | 16 | 3 | 5 | 79 |
| SCR Moving Bed | 10 | 5 | 7 | 14 | 9 | 3 | 14 | 16 | 3 | 5 | 86 |
| SCR Parallel Flow | 10 | 7 | 7 | 14 | 9 | 3 | 14 | 16 | 3 | 5 | 88 |
| Absorption - Oxidation | 7 | 3 | 6 | 3 | 4 | 2 | 8 | 13 | 3 | 5 | 54 |

TABLE A1.12. POINT VALUE SIMULTANEOUS $NO_x/SO_x$ PROCESS RATINGS: OIL-FIRED BOILERS - INTERMEDIATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Parallel Flow | 10 | 4 | 9 | 11 | 7 | 3 | 12 | 13 | 2 | 2 | 73 |
| Adsorption | * | | | | | | | | | | |
| Electron Beam Radiation | 4 | 5 | 3 | 6 | 4 | 2 | 8 | 7 | 3 | 5 | 47 |
| Absorption - Reduction | 8 | 3 | 7 | 9 | 5 | 4 | 7 | 10 | 3 | 2 | 58 |
| Oxidation - Absorption/Reduction | 8 | 4 | 6 | 4 | 2 | 2 | 8 | 16 | 3 | 2 | 55 |
| Oxidation - Absorption | 6 | 5 | 6 | 3 | 2 | 2 | 9 | 7 | 3 | 5 | 48 |

*Not Applicable - Does not meet removal requirements.

## TABLE A1.13. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: GAS-FIRED BOILERS - MODERATE CONTROL

| | Performance | Operation and maintenance | Environ-mental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Relia-bility | Status of development | Adapt-ability | Compati-bility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 12 | 7 | 8 | 14 | 10 | 4 | 14 | 16 | 3 | 5 | 93 |
| SCR Moving Bed | 12 | 5 | 8 | 14 | 10 | 4 | 14 | 16 | 3 | 5 | 91 |
| SCR Parallel Flow | 12 | 7 | 8 | 14 | 10 | 4 | 14 | 16 | 3 | 5 | 93 |
| Absorption - Oxidation | 8 | 4 | 7 | 2 | 6 | 2 | 8 | 9 | 3 | 5 | 58 |

## TABLE A1.14. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: GAS-FIRED BOILERS - STRINGENT CONTROL

| | Performance | Operation and maintenance | Environ-mental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Relia-bility | Status of development | Adapt-ability | Compati-bility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 8 | 7 | 8 | 8 | 10 | 4 | 14 | 16 | 3 | 5 | 83 |
| SCR Moving Bed | 8 | 5 | 8 | 8 | 10 | 4 | 14 | 16 | 3 | 5 | 81 |
| SCR Parallel Flow | 8 | 7 | 8 | 8 | 10 | 4 | 14 | 16 | 3 | 5 | 83 |
| Absorption - Oxidation | * | | | | | | | | | | |

*Not Applicable - Does not meet removal requirements.

## TABLE A1.15. POINT VALUE $NO_x$-ONLY PROCESS RATINGS: GAS-FIRED BOILERS - INTERMEDIATE CONTROL

| | Performance | Operation and maintenance | Environmental impact | Economic impact | Energy/ material impact | Boiler operation and safety | Reliability | Status of development | Adaptability | Compatibility | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| SCR Fixed Packed Bed | 10 | 7 | 8 | 14 | 10 | 4 | 14 | 16 | 3 | 5 | 87 |
| SCR Moving Bed | 10 | 5 | 8 | 14 | 10 | 4 | 14 | 16 | 3 | 5 | 85 |
| SCR Parallel Flow | 10 | 7 | 8 | 14 | 10 | 4 | 14 | 16 | 3 | 5 | 87 |
| Absorption - Oxidation | 8 | 4 | 7 | 2 | 6 | 2 | 8 | 9 | 3 | 5 | 58 |

# APPENDIX 2

# EXAMPLE OF TECHNIQUE FOR ECONOMIC SCALING

## SAMPLE CALCULATION

Most of the available economic data are for utility boilers. Various base capacities are utilized in the process developers' economic calculations. For the preliminary economic screening of alternative processes a base capacity of 20 MW was selected to represent industrial boilers. The capital cost figures were adjusted to a 20 MW cost by using the six-tenths rule.[1]

$$I = I_B \left(\frac{Q}{Q_B}\right)^{.6} \qquad \text{(Reference 1)}$$

where I = estimated 20 MW investment
$I_B$ = known base investment
Q = 20 MW
$Q_B$ = known base capacity

A sample calculation for the SCR-parallel passage (coal) process is shown below.

Size = 250 MW
Investment = \$4,000,000

$$I = (\$4{,}000{,}000)\left(\frac{20\ \text{MW}}{250\ \text{MW}}\right)^{.6}$$
$$= (\$4{,}000{,}000)\ (.220)$$
$$= \$879{,}000$$

$$\text{Capital Cost (\$/kW)} = \frac{\$879{,}000}{(20\ \text{MW})\left(\frac{1000\ \text{kW}}{\text{MW}}\right)}$$
$$= \$44/\text{kW}$$

The results of these calculations for all systems considered are contained in Tables A2.1 and A2.2.

Reference: 1. Rudd, D. F., C. C. Watson, Strategy of Process Engineering, 1968, p.121.

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Note: The six-tenths factor was used only during this preliminary phase to put reported costs for the various processes on a consistent basis. The six-tenths rule was not used in the development of process economics for this report.

TABLE A2.1. ECONOMICS - CAPITAL COST, $/kW

| | Coal | Oil | Gas |
|---|---|---|---|
| SCR fixed packed bed | 130* | 70 | 27 |
| SCR parallel flow | 44 | 39 | NA |
| SCR moving bed | 92* | 70 | NA |
| SCR parallel flow, $NO_x/SO_x$ | 475 | NA | NA |
| Adsorption | 215 | NA | NA |
| Electron Beam Radiation | 302 | NA | NA |
| Absorption - Reduction | 413 | 187 | NA |
| Absorption - Oxidation | NA | NA | NA |
| Oxidation - Absorption - Reduction | NA | 231 | NA |
| Oxidation - Absorption | NA | NA | 254 |

*Includes particulate removal

NA = Not Available

TABLE A2.2. ECONOMICS - OPERATING COST, MILLS/kWh

| | Coal | Oil | Gas |
|---|---|---|---|
| SCR fixed packed bed | 2.1 | 1.9 | 1.2 |
| SCR parallel flow | 1.5 | NA | NA |
| SCR moving bed | 2.0 | 1.8 | NA |
| SCR parallel flow, $NO_x/SO_x$ | 5.0 | NA | NA |
| Adsorption | 2.3 | NA | NA |
| Electron Beam Radiation | NA | NA | NA |
| Absorption - Reduction | 7.4 | 5.4 | NA |
| Absorption - Oxidation | NA | NA | NA |
| Oxidation - Absorption - Reduction | NA | 6.4 | NA |
| Oxidation - Absorption | NA | NA | NA |

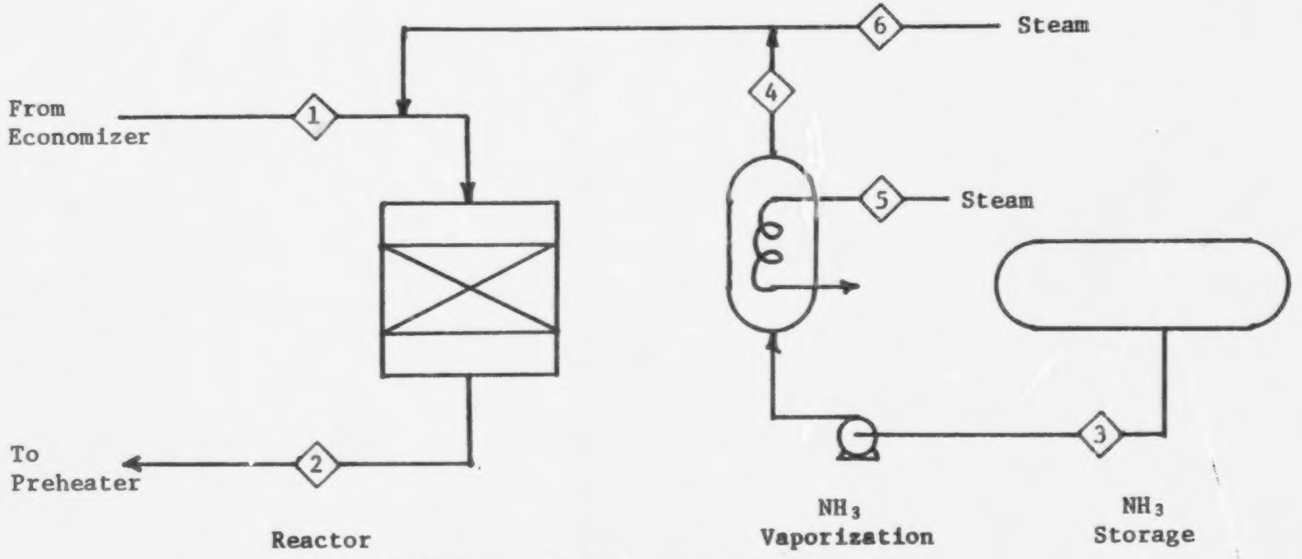
NA = Not Available

# APPENDIX 3

# MATERIAL BALANCES FOR COAL-FIRED BOILERS

MATERIAL BALANCE

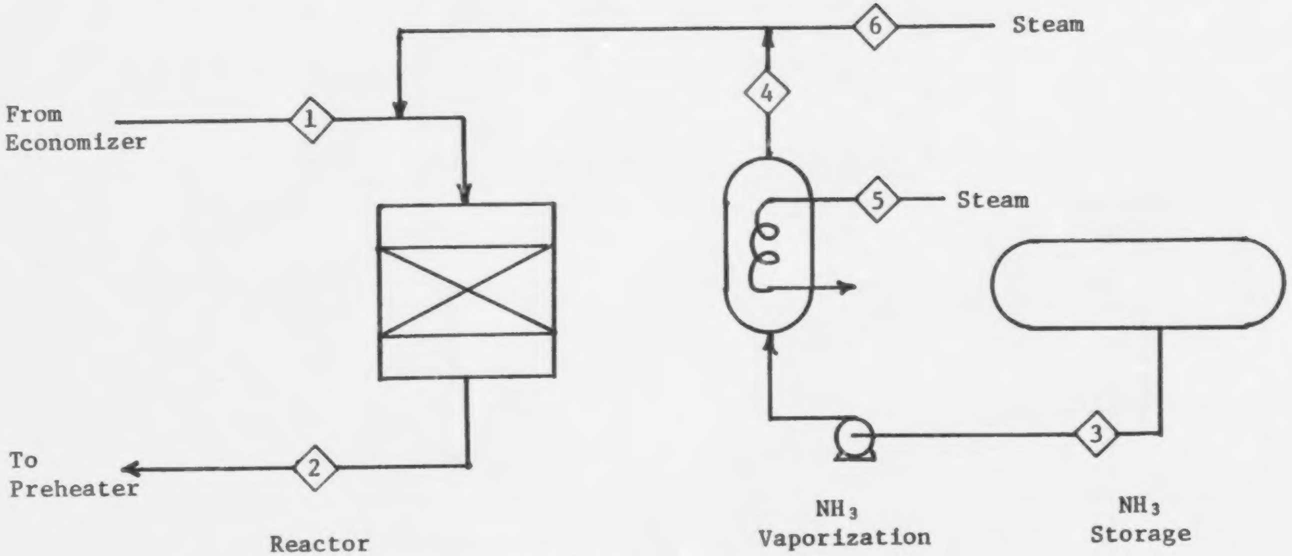
Underfeed Stoker
Parallel Flow SCR
Intermediate Control


Steam
From Economizer
1
6
4
5
Steam
3
To Preheater
2
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | ① | ② | ③ | ④ | ⑤ | ⑥ |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 98,040 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 431 | 431 | - | - | - | - |
| $CO_2$ | 67.9 | 67.9 | - | - | - | - |
| $H_2O$ | 49.9 | 51.2 | - | - | 0.10 | 0.94 |
| $O_2$ | 37.2 | 37.2 | - | - | - | - |
| $NO_x$ | 0.23 | 0.05 | - | - | - | - |
| $SO_x$ | 0.25 | 0.25 | - | - | - | - |
| $NH_3$ | - | <.01 | 0.19 | 0.19 | - | - |

## MATERIAL BALANCE

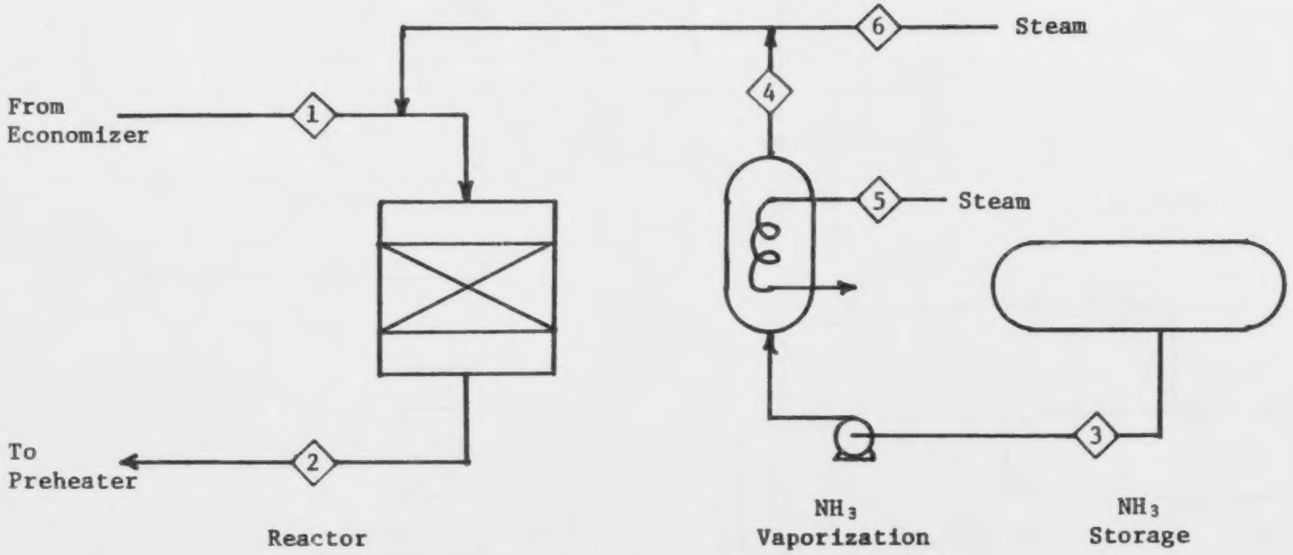
Chaingrate Stoker
Parallel Flow SCR
Stringent Control


6
Steam
From Economizer
1
4
5
Steam
To Preheater
2
3
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,770 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1081 | 1082 | - | - | - | - |
| $CO_2$ | 170 | 170 | - | - | - | - |
| $H_2O$ | 125 | 129 | - | - | 0.31 | 2.8 |
| $O_2$ | 93.2 | 93.1 | - | - | - | - |
| $NO_x$ | 0.58 | 0.06 | - | - | - | - |
| $SO_x$ | 0.63 | 0.63 | - | - | - | - |
| $NH_3$ | - | 0.01 | 0.55 | 0.55 | - | - |

## MATERIAL BALANCE

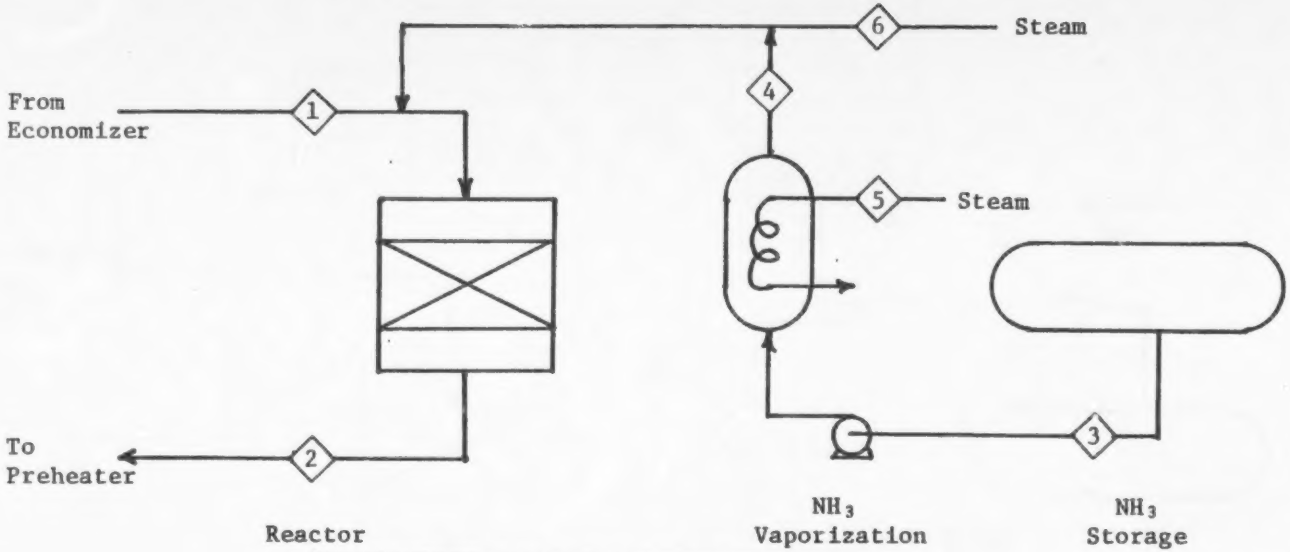
Chaingrate Stoker
Parallel Flow SCR
Intermediate Control


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1081 | 1082 | - | - | - | - |
| $CO_2$ | 170 | 170 | - | - | - | - |
| $H_2O$ | 125 | 128 | - | - | 0.26 | 2.4 |
| $O_2$ | 93.2 | 93.1 | - | - | - | - |
| $NO_x$ | 0.58 | 0.12 | - | - | - | - |
| $SO_x$ | 0.63 | 0.63 | - | - | - | - |
| $NH_3$ | - | <.01 | 0.48 | 0.48 | - | - |

MATERIAL BALANCE

Chaingrate Stoker
Parallel Flow SCR
Moderate Control

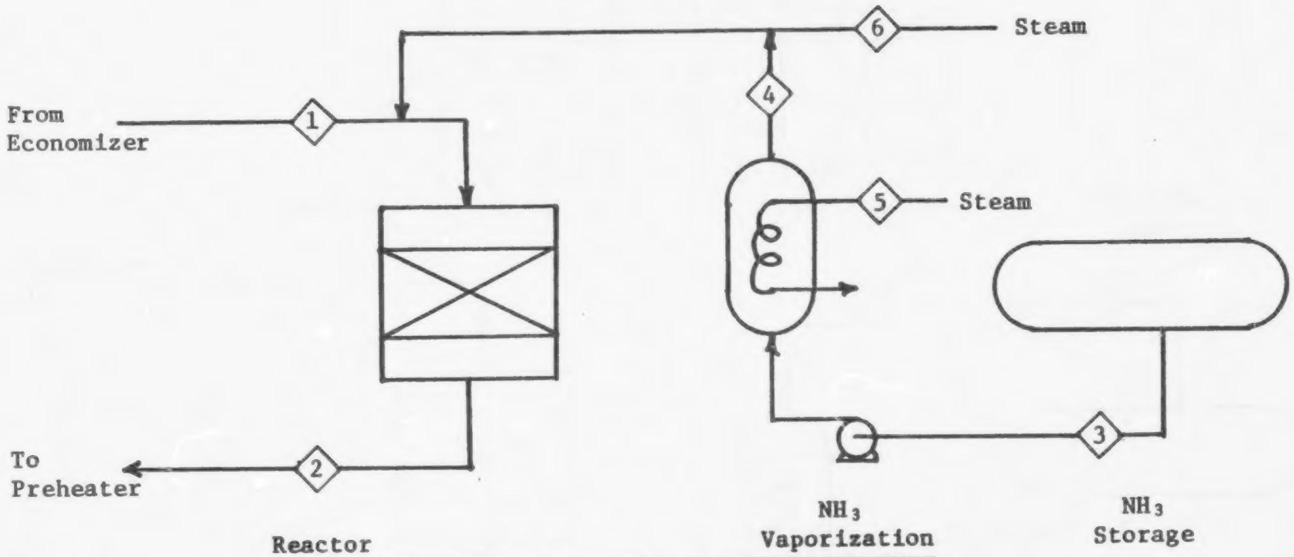


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
Reactor
NH₃ Vaporization
NH₃ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1081 | 1081 | - | - | - | - |
| $CO_2$ | 170 | 170 | - | - | - | - |
| $H_2O$ | 125 | 128 | - | - | 0.23 | 2.1 |
| $O_2$ | 93.2 | 93.1 | - | - | - | - |
| $NO_x$ | 0.58 | 0.17 | - | - | - | - |
| $SO_x$ | 0.63 | 0.63 | - | - | - | - |
| $NH_3$ | - | <.01 | 0.41 | 0.41 | - | - |

## MATERIAL BALANCE

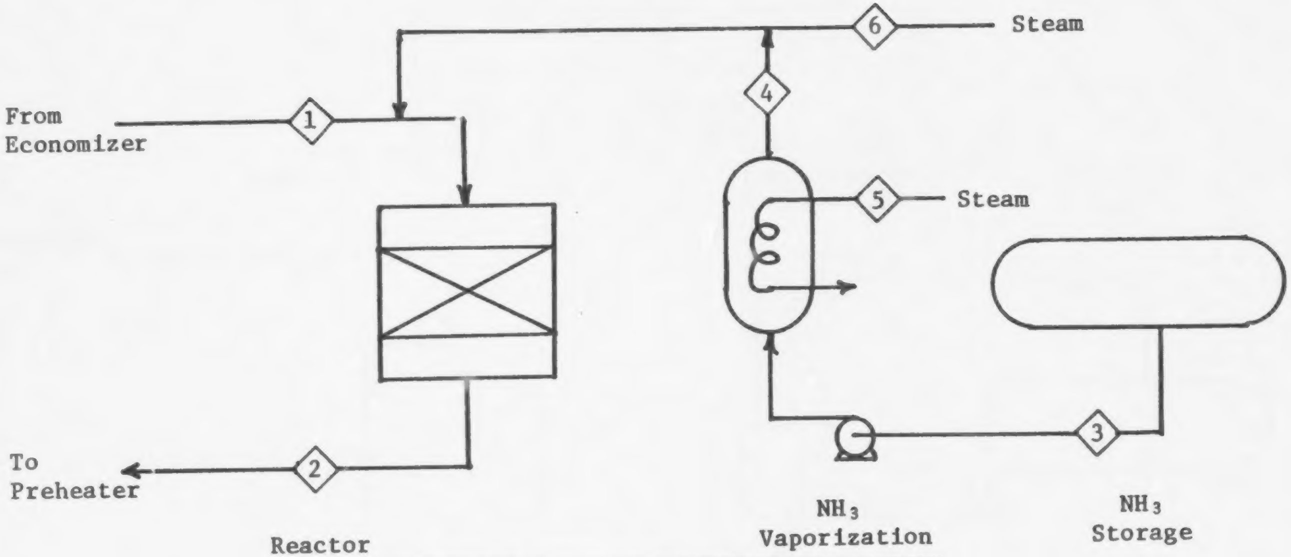
Spreader Stoker
Parallel Flow SCR
Intermediate Control


Steam
6
4
From
Economizer
1
5
Steam
To
Preheater
2
3
NH$_3$
Vaporization
NH$_3$
Storage
Reactor

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,810 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 2160 | 2161 | - | - | - | - |
| $CO_2$ | 339 | 339 | - | - | - | - |
| $H_2O$ | 250 | 256 | - | - | 0.52 | 4.7 |
| $O_2$ | 186 | 186 | - | - | - | - |
| $NO_x$ | 1.15 | 0.23 | - | - | - | - |
| $SO_x$ | 1.26 | 1.26 | - | - | - | - |
| $NH_3$ | - | 0.01 | 0.94 | 0.94 | - | - |

## MATERIAL BALANCE

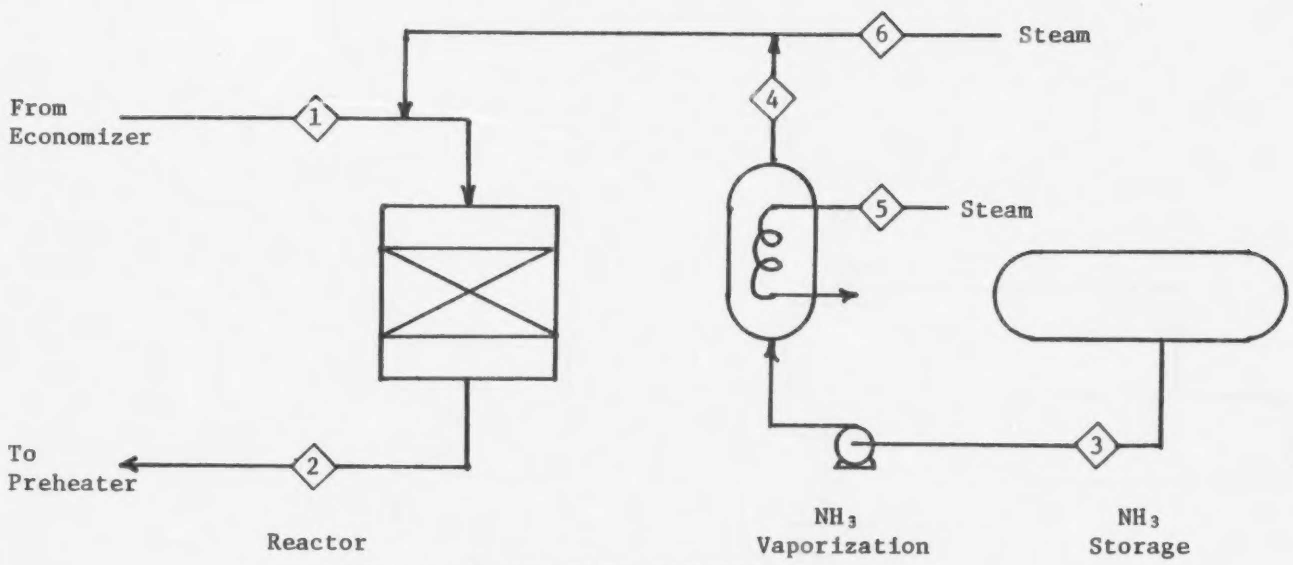
Pulverized Coal
Parallel Flow SCR
Stringent Control


Steam
From Economizer
1
6
4
Steam
5
To Preheater
2
3
Reactor
NH₃ Vaporization
NH₃ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,150 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 2495 | 2497 | - | - | - | - |
| $CO_2$ | 453 | 453 | - | - | - | - |
| $H_2O$ | 325 | 336 | - | - | 0.98 | 8.8 |
| $O_2$ | 149 | 149 | - | - | - | - |
| $NO_x$ | 1.86 | 0.19 | - | - | - | - |
| $SO_x$ | 1.69 | 1.69 | - | - | - | - |
| $NH_3$ | -- | 0.03 | 1.76 | 1.76 | - | - |

MATERIAL BALANCE

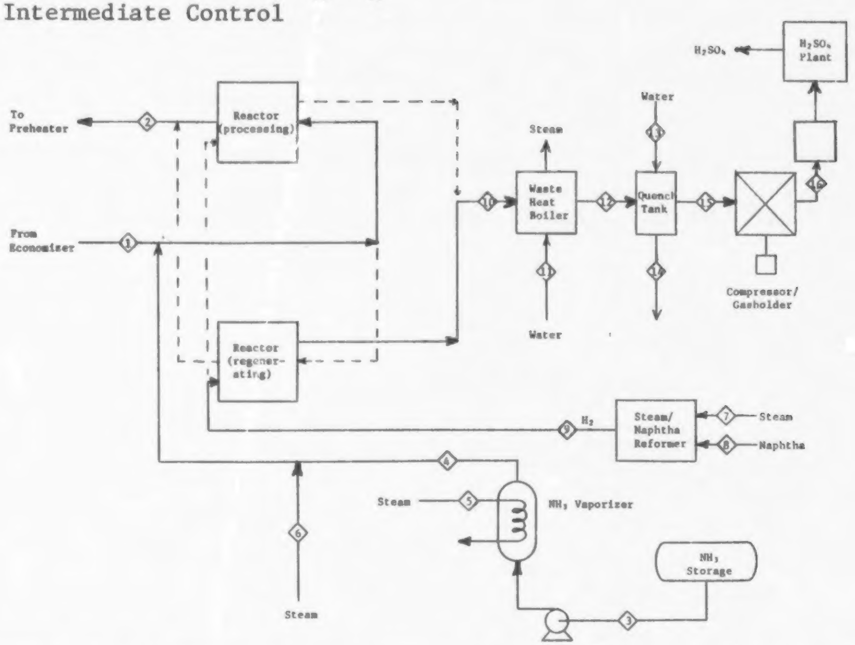
Pulverized Coal
Parallel Flow SCR
Moderate Control


From Economizer
1
2
To Preheater
Reactor
6
Steam
4
5
Steam
3
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 288 | 429 | 408 |
| P, Pa | 98,600 | 97,730 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 2495 | 2496 | - | - | - | - |
| $CO_2$ | 453 | 453 | - | - | - | - |
| $H_2O$ | 325 | 334 | - | - | 0.73 | 6.6 |
| $O_2$ | 149 | 149 | - | - | - | - |
| $NO_x$ | 1.86 | 0.56 | - | - | - | - |
| $SO_x$ | 1.69 | 1.69 | - | - | - | - |
| $NH_3$ | - | <.01 | 1.32 | 1.32 | - | - |

## MATERIAL BALANCE

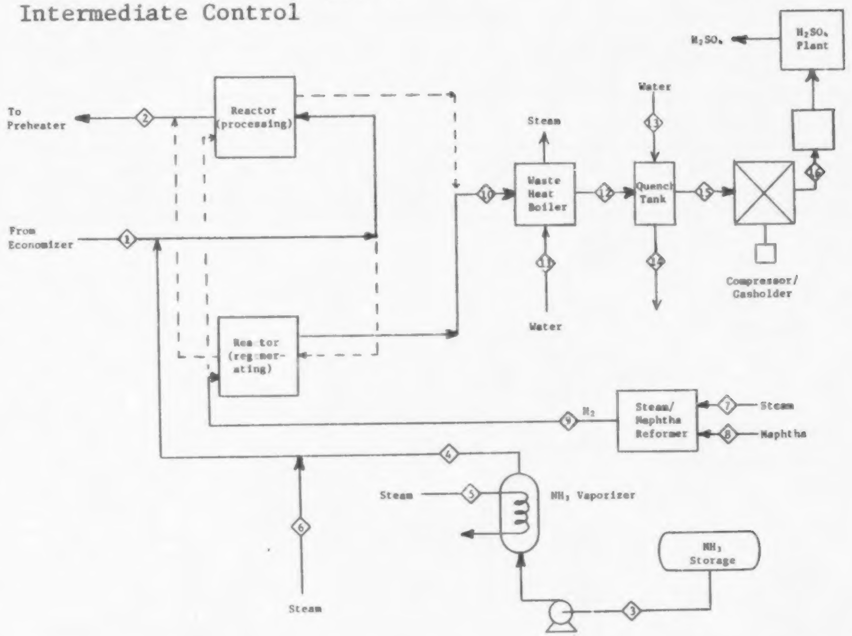
Underfeed Stoker (High Sulfur Eastern)
Parallel Flow SCR - $NO_x/SO_x$
Intermediate Control


To Preheater
2
Reactor (processing)
From Economizer
1
Reactor (regenerating)
10
Steam
Waste Heat Boiler
11
Water
12
Water
13
Quench Tank
14
15
Compressor/Gasholder
16
$H_2SO_4$ Plant
$H_2SO_4$
9
$H_2$
Steam/Naphtha Reformer
7
Steam
8
Naphtha
4
Steam
5
$NH_3$ Vaporizer
6
Steam
$NH_3$ Storage
3

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 | 789 | 666 | 704 | 700 | 411 | 450 | 300 | 346 | 389 | 389 |
| P, Pa | 98,600 | 96,150 | 615,000 | 752,000 | 552,000 | 310,000 | 665,000 | 860,000 | 274,000 | 177,000 | 345,000 | 170,000 | 276,000 | 276,000 | 165,000 | ~1.3x10^7 |
| kg-mole/hr | | | | | | | | | | | | | | | | |
| $N_2$ | 428 | 428 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $CO_2$ | 62.2 | 62.2 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2O$ | 42.2 | 43.6 | - | - | 0.14 | 1.18 | 7.20 | - | 2.52 | 9.71 | 2.93 | 9.71 | 7.54 | 7.54 | 9.71 | - |
| $O_2$ | 36.8 | 36.8 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $NO_x$ | 0.20 | 0.04 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $SO_x$ | 1.20 | 0.18 | - | - | - | - | - | - | - | 1.02 | - | 1.02 | - | <0.001 | 1.02 | 1.02 |
| $NH_3$ | - | 0.01 | 0.24 | 0.24 | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2$ | - | - | - | - | - | - | - | - | 7.19 | - | - | - | - | - | - | - |
| Naphtha | - | - | - | - | - | - | - | 0.34 | - | - | - | - | - | - | - | - |

## MATERIAL BALANCE

Underfeed Stoker (Low Sulfur Western)
Parallel Flow SCR - $NO_x/SO_x$
Intermediate Control

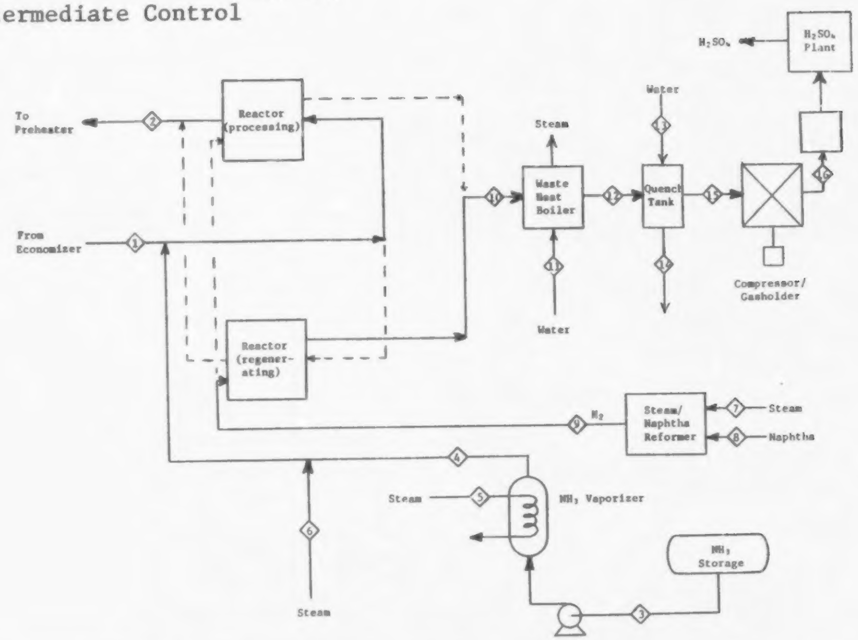


To Preheater
2
Reactor (processing)
From Economizer
1
Reactor (regener-ating)
10
Steam
Waste Heat Boiler
11
Water
12
Water
13
Quench Tank
14
15
Compressor/ Gasholder
16
$H_2SO_4$ Plant
$H_2SO_4$
9
$H_2$
Steam/ Naphtha Reformer
7
Steam
8
Naphtha
4
Steam
5
$NH_3$ Vaporizer
6
Steam
$NH_3$ Storage
3

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 | 789 | 666 | 704 | 700 | 411 | 450 | 300 | 346 | 389 | 389 |
| P, Pa | 98,600 | 96,150 | 615,000 | 752,000 | 552,000 | 310,000 | 665,000 | 860,000 | 274,000 | 177,000 | 345,000 | 170,000 | 276,000 | 276,000 | 165,000 | ~$1.3 \times 10^7$ |
| kg-mole/hr | | | | | | | | | | | | | | | | |
| $N_2$ | 431 | 431 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $CO_2$ | 67.9 | 67.9 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2O$ | 49.9 | 51.6 | - | - | 0.16 | 1.38 | 1.51 | - | 0.53 | 2.03 | 0.63 | 2.03 | 1.59 | 1.59 | 2.03 | - |
| $O_2$ | 37.2 | 37.2 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $NO_x$ | 0.23 | 0.05 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $SO_x$ | 0.25 | 0.04 | - | - | - | - | - | - | - | 0.21 | - | 0.21 | - | <0.0001 | 0.21 | 0.21 |
| $NH_3$ | - | 0.01 | 0.28 | 0.28 | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2$ | - | - | - | - | - | - | - | - | 1.50 | - | - | - | - | - | - | - |
| Naphtha | - | - | - | - | - | - | - | 0.07 | - | - | - | - | - | - | - | - |

MATERIAL BALANCE

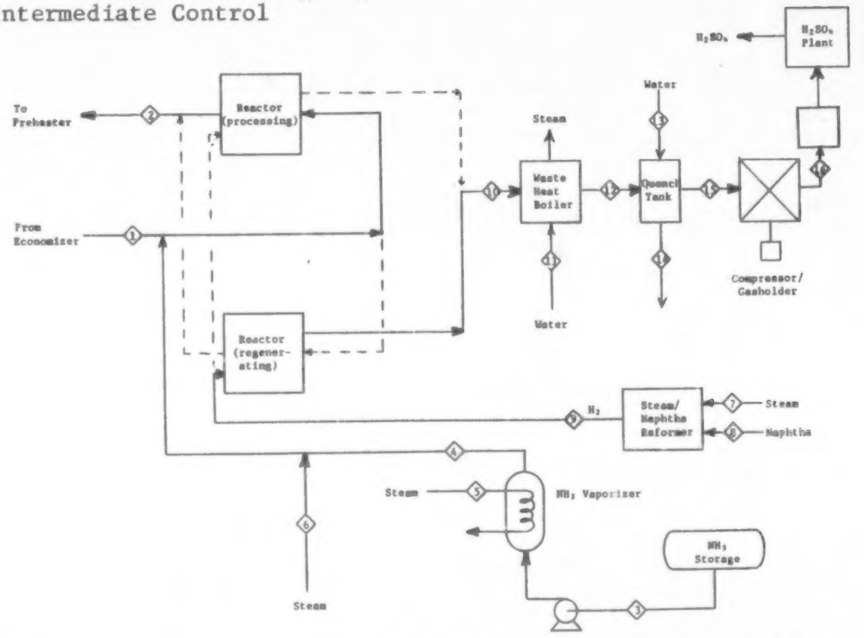
Pulverized Coal (High Sulfur Eastern)
Parallel Flow SCR - $NO_x/SO_x$
Intermediate Control


To Preheater
From Economizer
Reactor (processing)
Reactor (regenerating)
Waste Heat Boiler
Quench Tank
Compressor/ Gasholder
$H_2SO_4$ Plant
$H_2SO_4$
Steam
Water
Water
Steam/ Naphtha Reformer
$H_2$
Steam
Naphtha
$NH_3$ Vaporizer
$NH_3$ Storage
Steam
Steam
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16

| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 | 789 | 666 | 704 | 700 | 411 | 450 | 300 | 346 | 389 | 389 |
| P, Pa | 98,600 | 96,150 | 6[illegible] | [illegible],000 | 552,000 | 310,000 | 665,000 | 860,000 | 274,000 | 177,000 | 345,000 | 170,000 | 276,000 | 276,000 | 165,000 | ~1.3x10$^7$ |
| kg-mole/hr | | | | | | | | | | | | | | | | |
| $N_2$ | 2470 | 2470 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $CO_2$ | 415 | 415 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2O$ | 271 | 282 | - | - | 1.11 | 9.40 | 48.0 | - | 16.8 | 64.7 | 19.6 | 64.7 | 50.4 | 50.4 | 64.7 | - |
| $O_2$ | 148 | 148 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $NO_x$ | 1.57 | 0.31 | - | - | - | - | - | - | - | - | - | - | - | - | | - |
| $SO_x$ | 7.99 | 1.20 | - | - | - | - | - | - | - | 6.79 | - | 6.79 | - | <0.01 | 6.79 | 6.79 |
| $NH_3$ | - | 0.07 | 1.88 | 1.88 | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2$ | - | - | - | - | - | - | - | - | 47.9 | - | - | - | - | - | - | - |
| Naphtha | - | - | - | - | - | - | - | 2.23 | - | - | - | - | - | - | - | - |

## MATERIAL BALANCE

Pulverized Coal (Low Sulfur Western)
Parallel Flow SCR - $NO_x/SO_x$
Intermediate Control


To Preheater
From Economizer
Reactor (processing)
Reactor (regenerating)
Steam
Waste Heat Boiler
Water
Quench Tank
Compressor/ Gasholder
$H_2SO_4$ Plant
$H_2SO_4$
Steam/ Naphtha Reformer
$N_2$
Naphtha
$NH_3$ Vaporizer
$NH_3$ Storage
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16

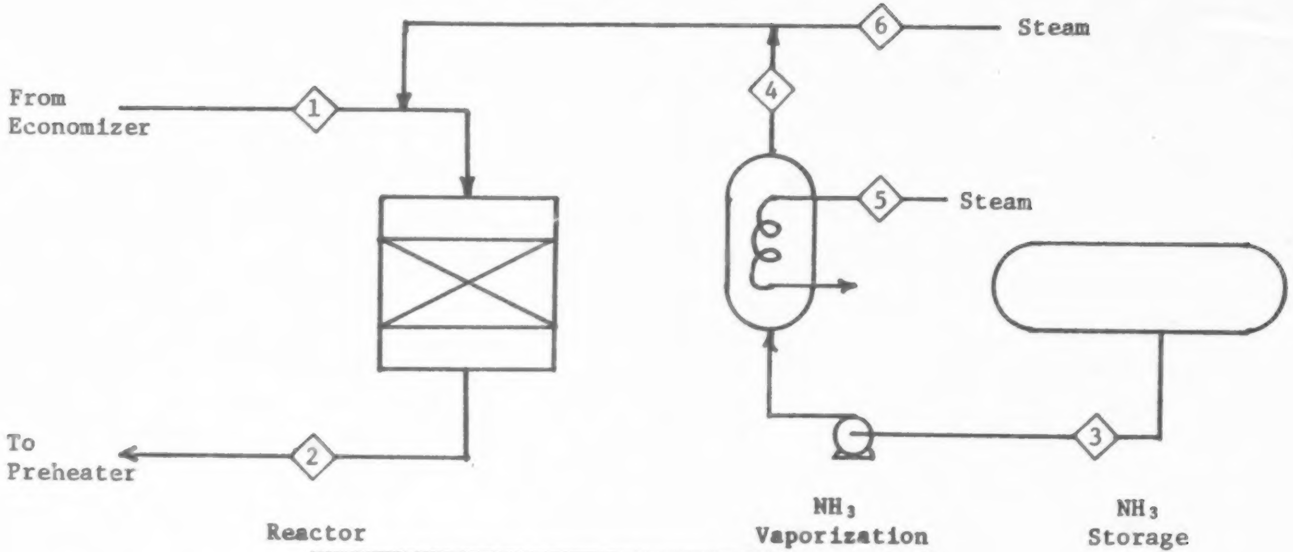
|  | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 | 789 | 666 | 704 | 700 | 411 | 450 | 300 | 346 | 389 | 389 |
| P, Pa | 98,600 | 96,150 | 615,000 | 752,000 | 552,000 | 310,000 | 665,000 | 860,000 | 274,000 | 177,000 | 345,000 | 170,000 | 276,000 | 276,000 | 165,000 | ~$1.3 \times 10^7$ |
| kg-mole/hr |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| $N_2$ | 2490 | 2490 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $CO_2$ | 453 | 453 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2O$ | 325 | 340 | - | - | 1.32 | 11.2 | 10.2 | - | 3.55 | 13.7 | 4.14 | 13.7 | 10.6 | 10.6 | 13.7 | - |
| $O_2$ | 149 | 149 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $NO_x$ | 1.86 | 0.37 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $SO_x$ | 1.69 | 0.25 | - | - | - | - | - | - | - | 1.44 | - | 1.44 | - | <0.001 | 1.44 | 1.44 |
| $NH_3$ | - | 0.07 | 2.23 | 2.23 | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2$ | - | - | - | - | - | - | - | - | 10.1 | - | - | - | - | - | - | - |
| Naphtha | - | - | - | - | - | - | - | 0.47 | - | - | - | - | - | - | - | - |

# APPENDIX 4

# MATERIAL BALANCES FOR OIL-FIRED BOILERS

## MATERIAL BALANCE

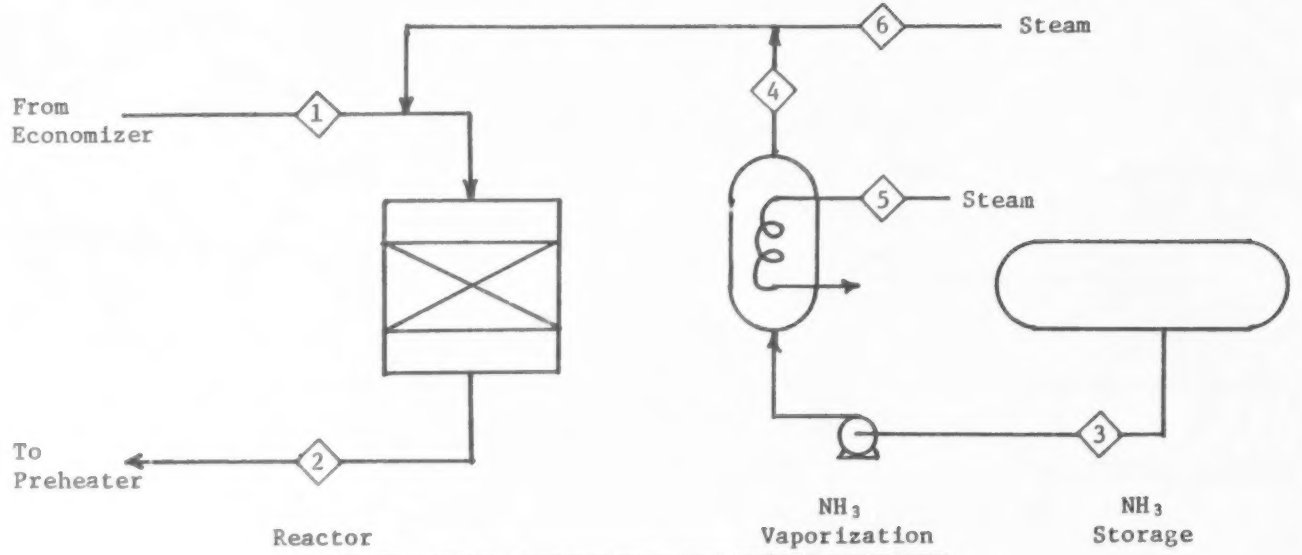
Distillate Oil (4.4 MW)
Fixed Packed Bed SCR
Stringent Control


From Economizer
To Preheater
Steam
Steam
1
2
3
4
5
6
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,420 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 168 | 168 | - | - | - | - |
| $CO_2$ | 26.3 | 26.3 | - | - | - | - |
| $H_2O$ | 28.6 | 28.7 | - | - | 0.011 | 0.101 |
| $O_2$ | 5.7 | 5.7 | - | - | - | - |
| $NO_x$ | 0.023 | 0.002 | - | - | - | - |
| $SO_x$ | 0.054 | 0.054 | - | - | - | - |
| $NH_3$ | - | <.001 | 0.020 | 0.020 | - | - |

MATERIAL BALANCE

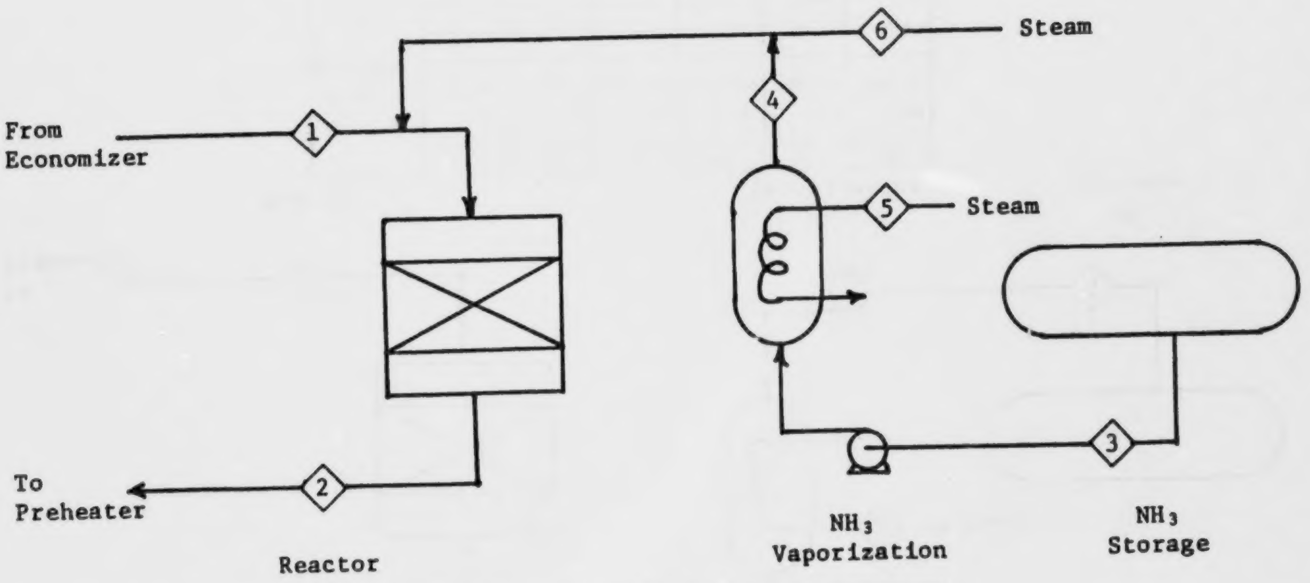
Distillate Oil (4.4 MW)
Fixed Packed Bed SCR
Moderate Control


Steam
6
From
Economizer
1
4
5
Steam
To
Preheater
2
3
$NH_3$
Vaporization
$NH_3$
Storage
Reactor

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,890 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 168 | 168 | - | - | - | - |
| $CO_2$ | 26.3 | 26.3 | - | - | - | - |
| $H_2O$ | 28.6 | 28.7 | - | - | 0.009 | 0.078 |
| $O_2$ | 5.7 | 5.7 | - | - | - | - |
| $NO_x$ | 0.023 | 0.007 | - | - | - | - |
| $SO_x$ | 0.054 | 0.054 | - | - | - | - |
| $NH_3$ | - | <.001 | 0.016 | 0.016 | - | - |

## MATERIAL BALANCE

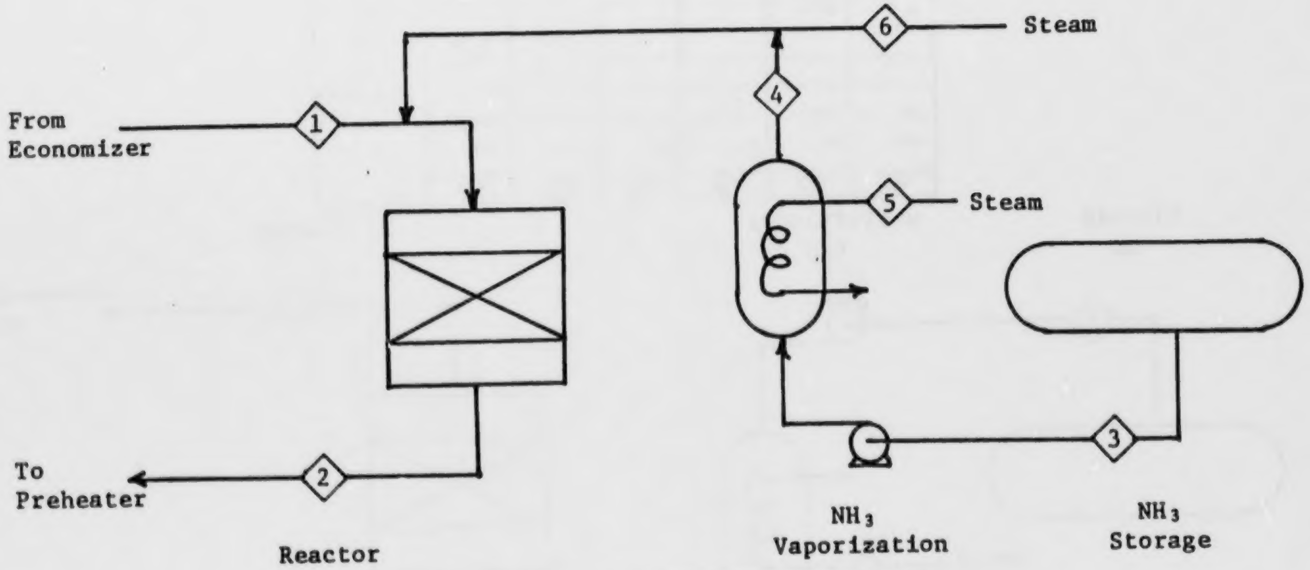
Distillate Oil (44 MW)
Fixed Packed Bed SCR
Stringent Control


Steam
From Economizer
1
6
4
Steam
5
To Preheater
2
3
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K<br>P, Pa | | | | | | |
| $N_2$ | 1470 | 1470 | - | - | - | - |
| $CO_2$ | 230 | 230 | - | - | - | - |
| $H_2O$ | 250 | 252 | - | - | 0.132 | 1.19 |
| $O_2$ | 50.0 | 50.0 | - | - | - | - |
| $NO_x$ | 0.24 | 0.02 | - | - | - | - |
| $SO_x$ | 0.54 | 0.54 | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.211 | 0.211 | - | - |

MATERIAL BALANCE

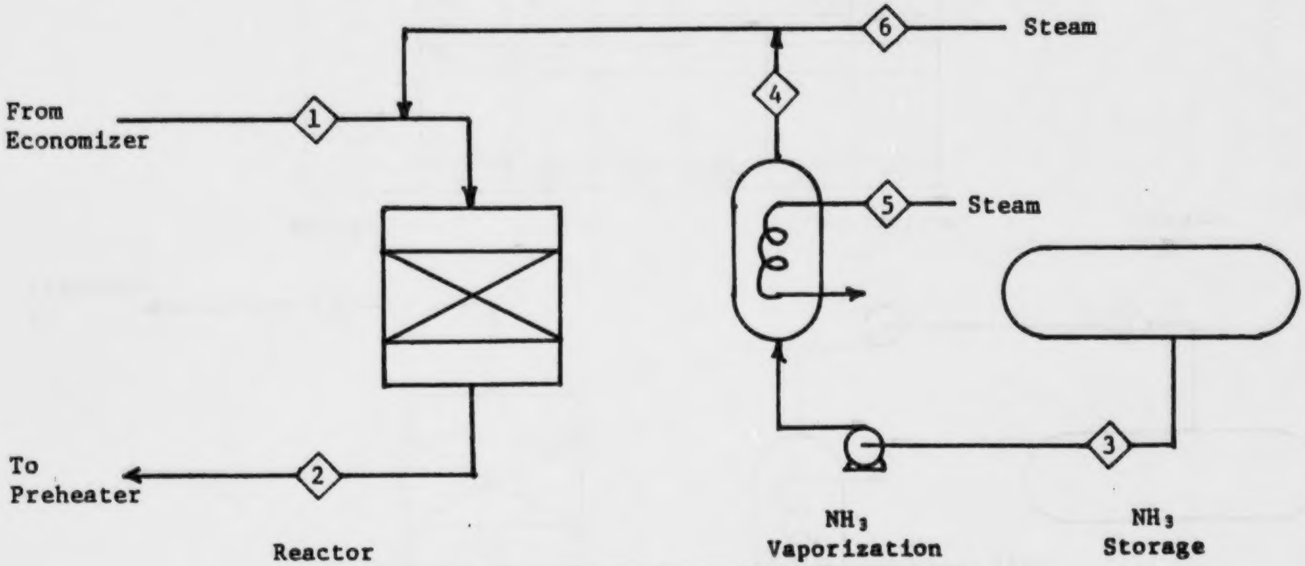
Distillate Oil (44 MW)
Fixed Packed Bed SCR
Moderate Control


Steam
Steam
From Economizer
To Preheater
Reactor
$NH_3$ Vaporization
$NH_3$ Storage
1
2
3
4
5
6

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | | | | | | |
| P, Pa | | | | | | |
| $N_2$ | 1470 | 1470 | - | - | - | - |
| $CO_2$ | 230 | 230 | - | - | - | - |
| $H_2O$ | 250 | 252 | - | - | 0.103 | 0.923 |
| $O_2$ | 50.0 | 50.0 | - | - | - | - |
| $NO_x$ | 0.24 | 0.08 | - | - | - | - |
| $SO_x$ | 0.54 | 0.54 | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.164 | 0.164 | - | - |

## MATERIAL BALANCE

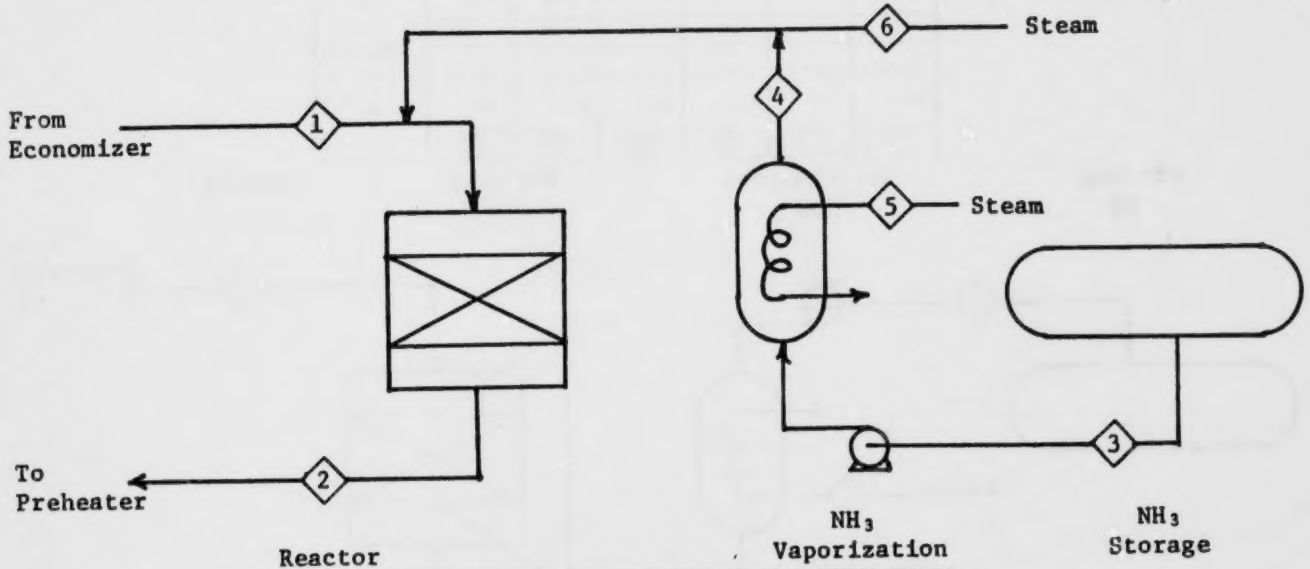
Residual Oil (8.8 MW)
Parallel Flow SCR
Stringent Control


Steam
From Economizer
1
6
4
Steam
5
To Preheater
2
3
Reactor
NH$_3$ Vaporization
NH$_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,890 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 297 | 297 | - | - | - | - |
| $CO_2$ | 49.8 | 49.8 | - | - | - | - |
| $H_2O$ | 44.2 | 45.2 | - | - | 0.09 | 0.78 |
| $O_2$ | 10.2 | 10.2 | - | - | - | - |
| $NO_x$ | 0.17 | 0.02 | - | - | - | - |
| $SO_x$ | 0.67 | 0.67 | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.16 | 0.16 | - | - |

MATERIAL BALANCE

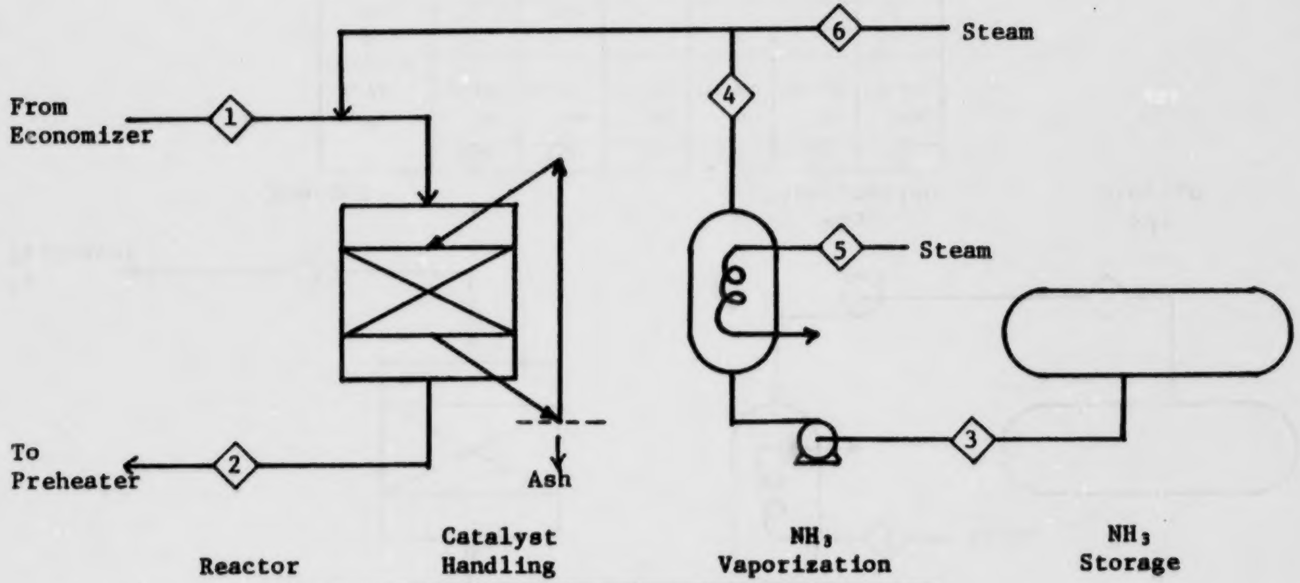
Residual Oil (8.8 MW)
Parallel Flow SCR
Moderate Control


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
$NH_3$ Vaporization
$NH_3$ Storage
Reactor

|  | ① | ② | ③ | ④ | ⑤ | ⑥ |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 98,180 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr |  |  |  |  |  |  |
| $N_2$ | 297 | 297 | - | - | - | - |
| $CO_2$ | 49.8 | 49.8 | - | - | - | - |
| $H_2O$ | 44.2 | 45.0 | - | - | 0.07 | 0.60 |
| $O_2$ | 10.2 | 10.2 | - | - | - | - |
| $NO_x$ | 0.17 | 0.05 | - | - | - | - |
| $SO_x$ | 0.67 | 0.67 | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.12 | 0.12 | - | - |

MATERIAL BALANCE

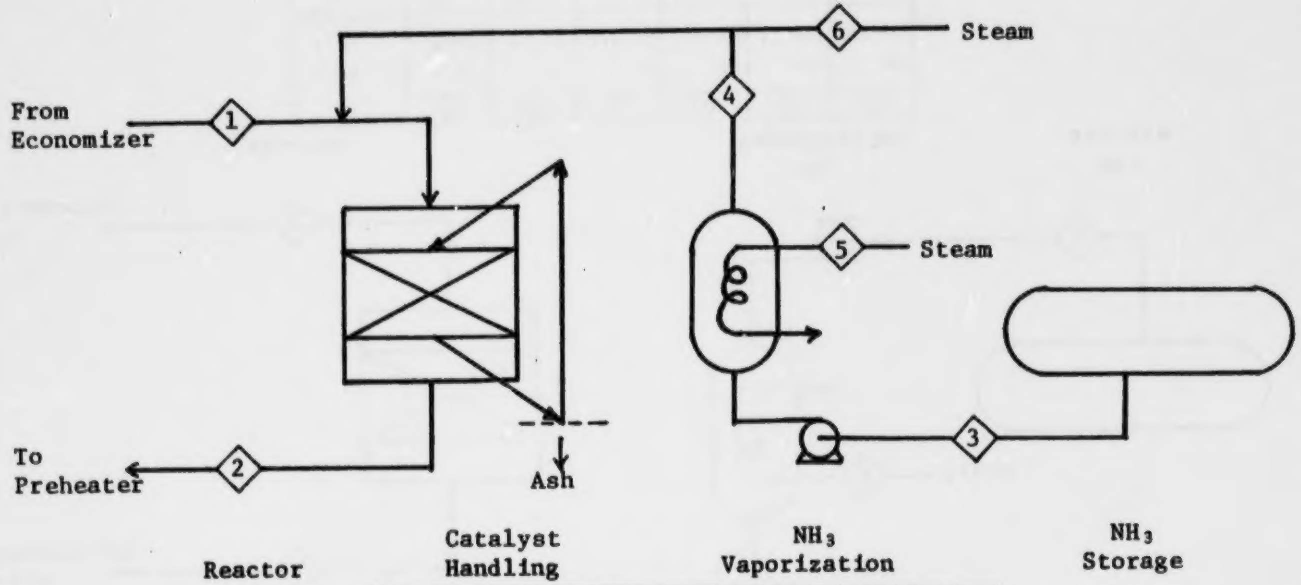
Residual Oil (8.8 MW)
Moving Bed SCR
Stringent Control


From Economizer
1
6
Steam
4
5
Steam
To Preheater
2
Ash
3
Reactor
Catalyst Handling
NH₃ Vaporization
NH₃ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 98,160 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 297 | 297 | - | - | - | - |
| $CO_2$ | 49.8 | 49.8 | - | - | - | - |
| $H_2O$ | 44.2 | 45.2 | - | - | 0.09 | 0.78 |
| $O_2$ | 10.2 | 10.2 | - | - | - | - |
| $NO_x$ | 0.17 | 0.02 | - | - | - | - |
| $SO_x$ | 0.67 | 0.67 | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.16 | 0.16 | - | - |

MATERIAL BALANCE

Residual Oil (8.8 MW)
Moving Bed SCR
Moderate Control

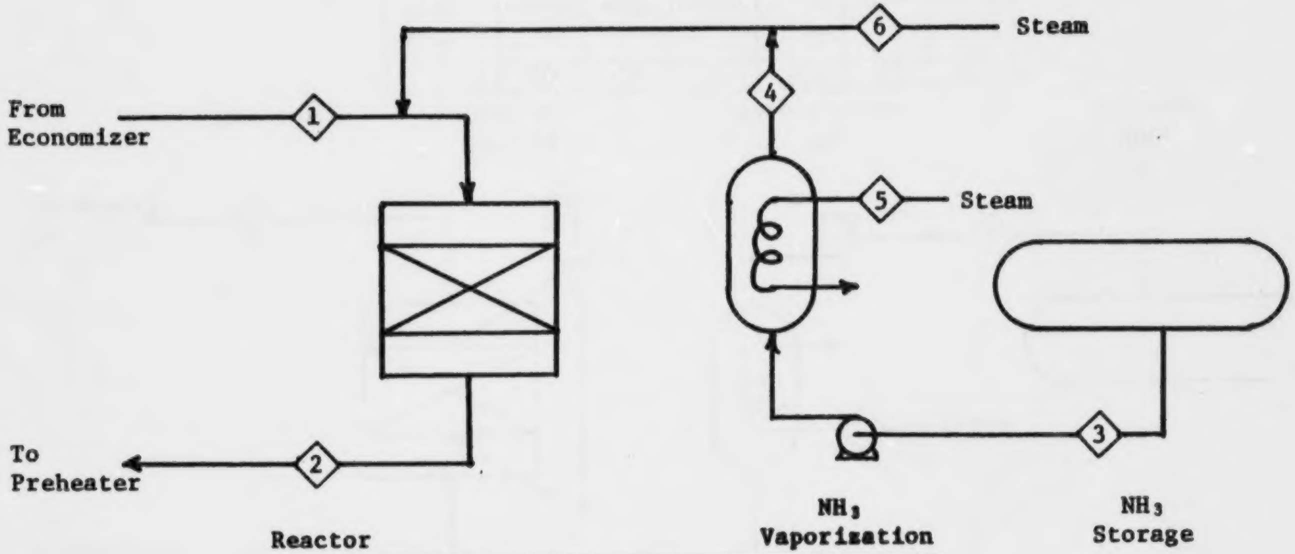


Steam
6
4
From Economizer
1
5
Steam
To Preheater
2
Ash
3
Reactor
Catalyst Handling
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 98,340 | 615,000 | 752,000 | 552,000 | 310,000 |
| $N_2$ | 297 | 297 | - | - | - | - |
| $CO_2$ | 49.8 | 49.8 | - | - | - | - |
| $H_2O$ | 44.2 | 45.0 | - | - | 0.07 | 0.60 |
| $O_2$ | 10.2 | 10.2 | - | - | - | - |
| $NO_x$ | 0.17 | 0.05 | - | - | - | - |
| $SO_x$ | 0.67 | 0.67 | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.12 | 0.12 | | |

## MATERIAL BALANCE

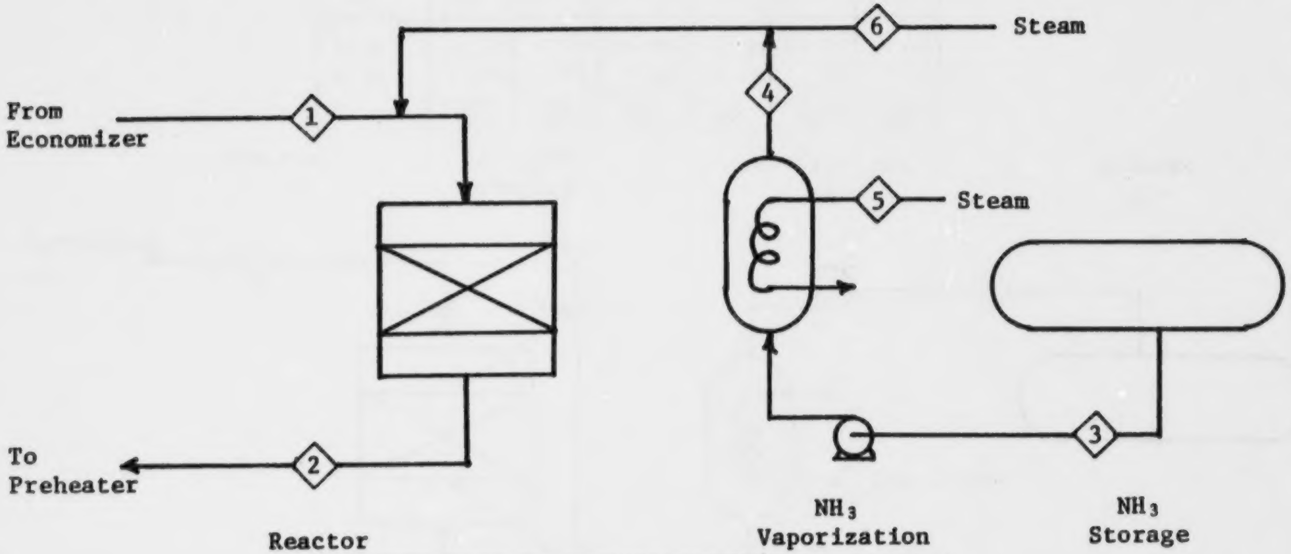
Residual Oil (44 MW)
Parallel Flow SCR
Stringent Control


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
Reactor
NH$_3$ Vaporization
NH$_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,460 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1489 | 1490 | - | - | - | - |
| $CO_2$ | 250 | 250 | - | - | - | - |
| $H_2O$ | 222 | 226 | - | - | 0.31 | 2.8 |
| $O_2$ | 51.1 | 51.0 | - | - | - | - |
| $NO_x$ | 0.59 | 0.06 | - | - | - | - |
| $SO_x$ | 3.34 | 3.34 | - | - | - | - |
| $NH_3$ | - | 0.02 | 0.56 | 0.56 | - | - |

MATERIAL BALANCE

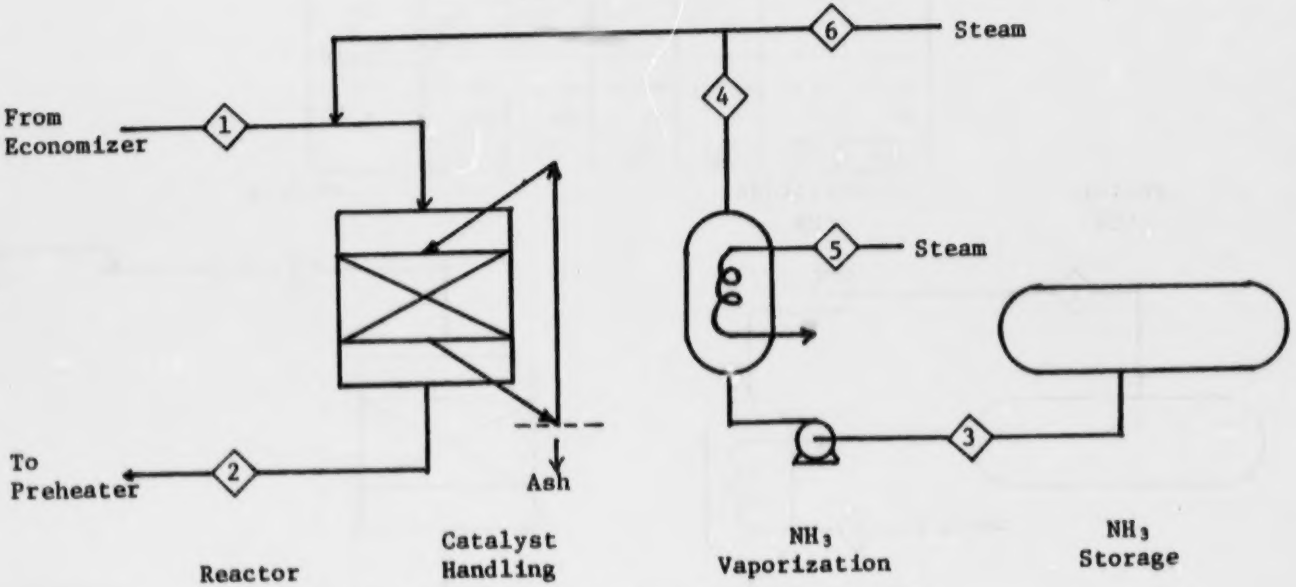
Residual Oil (44 MW)
Parallel Flow SCR
Moderate Control


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
NH₃ Vaporization
NH₃ Storage
Reactor

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,910 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1489 | 1489 | - | - | - | - |
| $CO_2$ | 250 | 250 | - | - | - | - |
| $H_2O$ | 222 | 225 | - | - | 0.19 | 2.1 |
| $O_2$ | 51.1 | 51.0 | - | - | - | - |
| $NO_x$ | 0.59 | 0.18 | - | - | - | - |
| $SO_x$ | 3.34 | 3.34 | - | - | - | - |
| $NH_3$ | - | <.01 | 0.42 | 0.42 | - | - |

## MATERIAL BALANCE

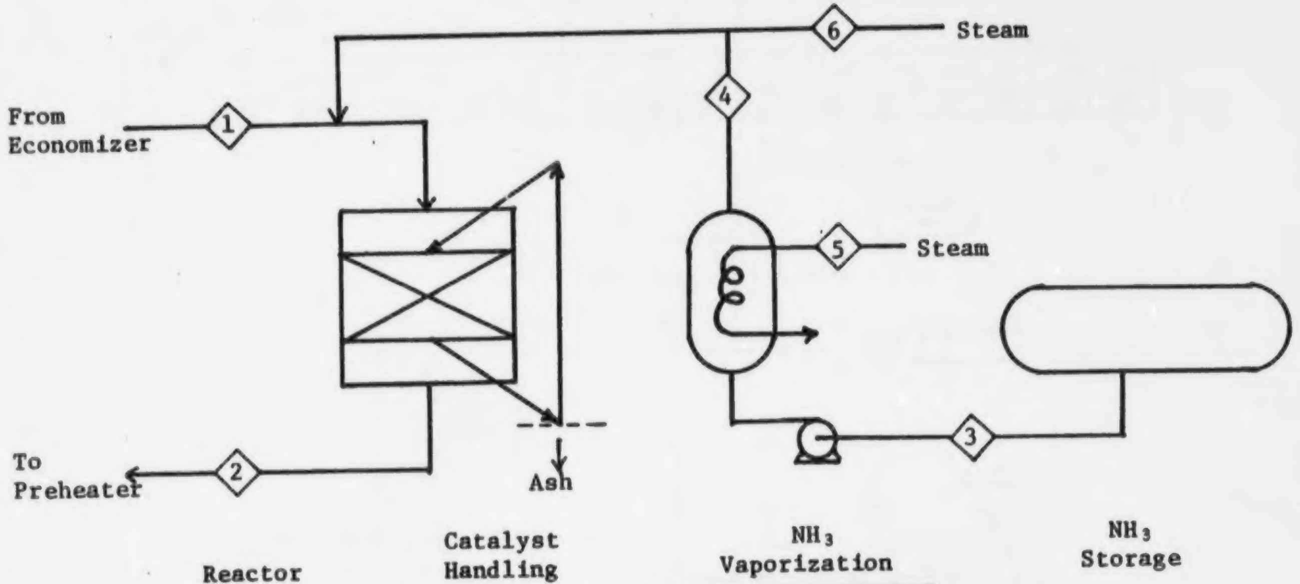
Residual Oil (44 MW)
Moving Bed SCR
Stringent Control


Steam
From Economizer
1
6
4
Steam
5
To Preheater
2
Ash
3
Reactor
Catalyst Handling
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,850 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1489 | 1490 | - | - | - | - |
| $CO_2$ | 250 | 250 | - | - | - | - |
| $H_2O$ | 222 | 226 | - | - | 0.31 | 2.8 |
| $O_2$ | 51.1 | 51.0 | - | - | - | - |
| $NO_x$ | 0.59 | 0.06 | - | - | - | - |
| $SO_x$ | 3.34 | 3.34 | - | - | - | - |
| $NH_3$ | - | 0.02 | 0.56 | 0.56 | - | - |

## MATERIAL BALANCE

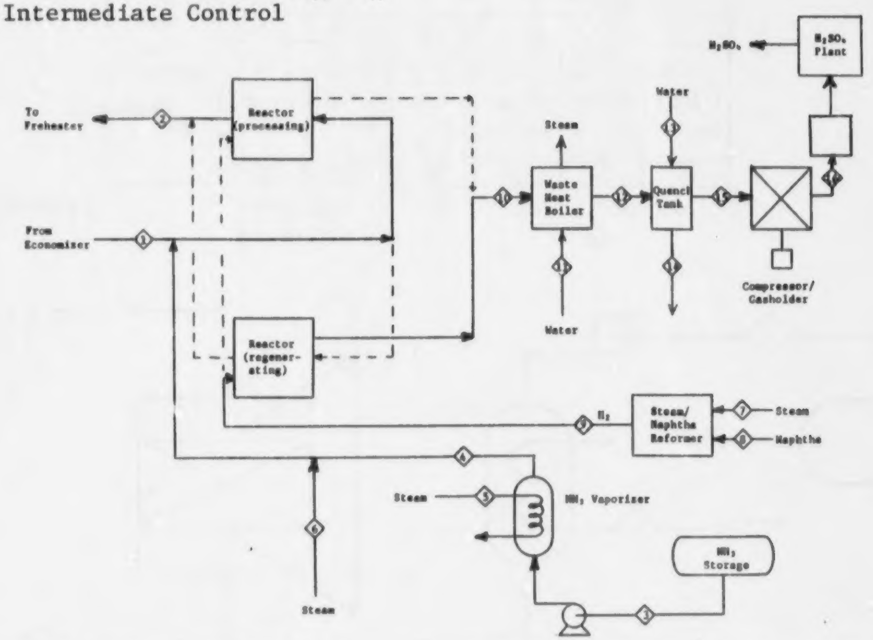
Residual Oil (44 MW)
Moving Bed SCR
Moderate Control


Steam
From Economizer
1
6
4
5
Steam
2
3
To Preheater
Ash
Reactor
Catalyst Handling
NH₃ Vaporization
NH₃ Storage

| | ① | ② | ③ | ④ | ⑤ | ⑥ |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 98,150 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 1489 | 1489 | - | - | - | - |
| $CO_2$ | 250 | 250 | - | - | - | - |
| $H_2O$ | 222 | 225 | - | - | 0.19 | 2.1 |
| $O_2$ | 51.1 | 51.0 | - | - | - | - |
| $NO_x$ | 0.59 | 0.18 | - | - | - | - |
| $SO_x$ | 3.34 | 3.34 | - | - | - | - |
| $NH_3$ | - | <.01 | 0.42 | 0.42 | - | - |

MATERIAL BALANCE

Residual Oil (44 MW)
Parallel Flow SCR - $NO_x/SO_x$
Intermediate Control



To Preheater
From Economizer
Reactor (processing)
Reactor (regenerating)
Waste Heat Boiler
Steam
Water
Quench Tank
Water
Compressor/ Gasholder
$H_2SO_4$ Plant
$H_2SO_4$
Steam/ Naphtha Reformer
Steam
Naphtha
$H_2$
$NH_3$ Vaporizer
Steam
$NH_3$ Storage
Steam
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16

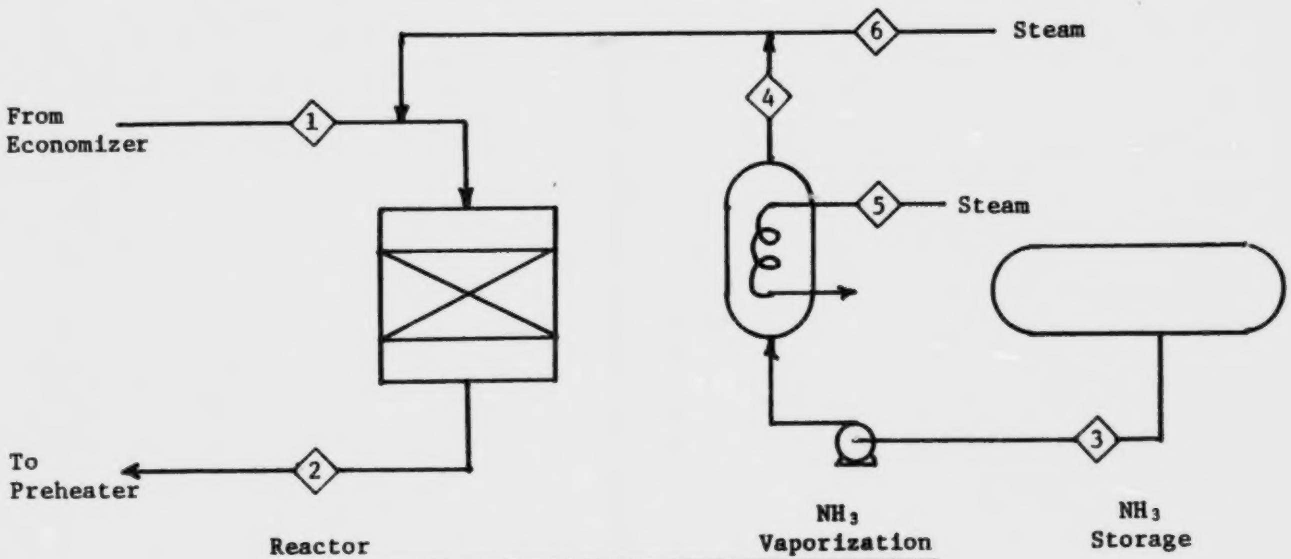
| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 | 789 | 666 | 704 | 700 | 411 | 450 | 300 | 346 | 389 | 389 |
| P, Pa | 98,600 | 96,150 | 615,000 | 752,000 | 552,000 | 310,000 | 665,000 | 860,000 | 274,000 | 177,000 | 345,000 | 170,000 | 276,000 | 276,000 | 165,000 | ~$1.3 \times 10^7$ |
| kg-mole/hr | | | | | | | | | | | | | | | | |
| $N_2$ | 1490 | 1490 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $CO_2$ | 250 | 250 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2O$ | 222 | 226 | - | - | 0.42 | 3.54 | 20.1 | - | 7.01 | 27.0 | 8.17 | 27.0 | 21.0 | 21.0 | 27.0 | - |
| $O_2$ | 51.1 | 51.1 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $NO_x$ | 0.59 | 0.12 | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| $SO_x$ | 3.34 | 0.50 | - | - | - | - | - | - | - | 2.84 | - | 2.84 | - | <0.01 | 2.84 | 2.84 |
| $NH_3$ | - | 0.04 | 0.71 | 0.71 | - | - | - | - | - | - | - | - | - | - | - | - |
| $H_2$ | - | - | - | - | - | - | - | - | 20.0 | - | - | - | - | - | - | - |
| Naphtha | - | - | - | - | - | - | - | 0.93 | - | - | - | - | - | - | - | - |

# APPENDIX 5

# MATERIAL BALANCES FOR NATURAL GAS-FIRED BOILERS

## MATERIAL BALANCE

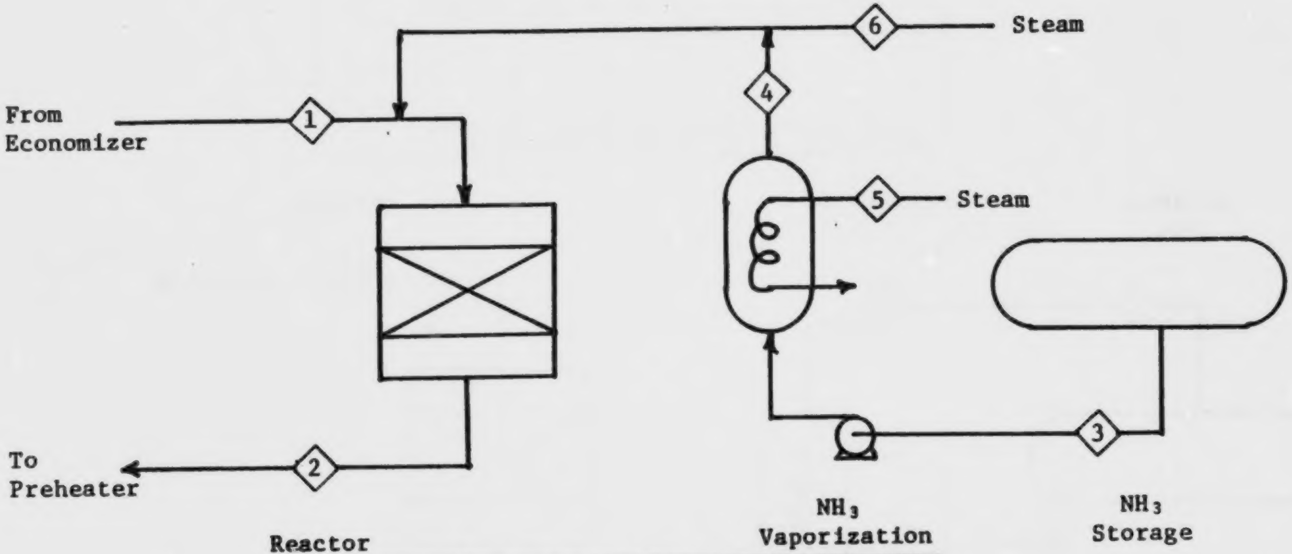
Natural Gas (4.4 MW)
Fixed Packed Bed SCR
Stringent Control


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,350 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 183 | 183 | - | - | - | - |
| $CO_2$ | 21.4 | 21.4 | - | - | - | - |
| $H_2O$ | 46.6 | 46.7 | - | - | 0.013 | 0.11 |
| $O_2$ | 6.3 | 6.3 | - | - | - | - |
| $NO_x$ | 0.026 | 0.003 | - | - | - | - |
| $SO_x$ | tr | tr | - | - | - | - |
| $NH_3$ | - | <.001 | 0.023 | 0.023 | - | - |

## MATERIAL BALANCE

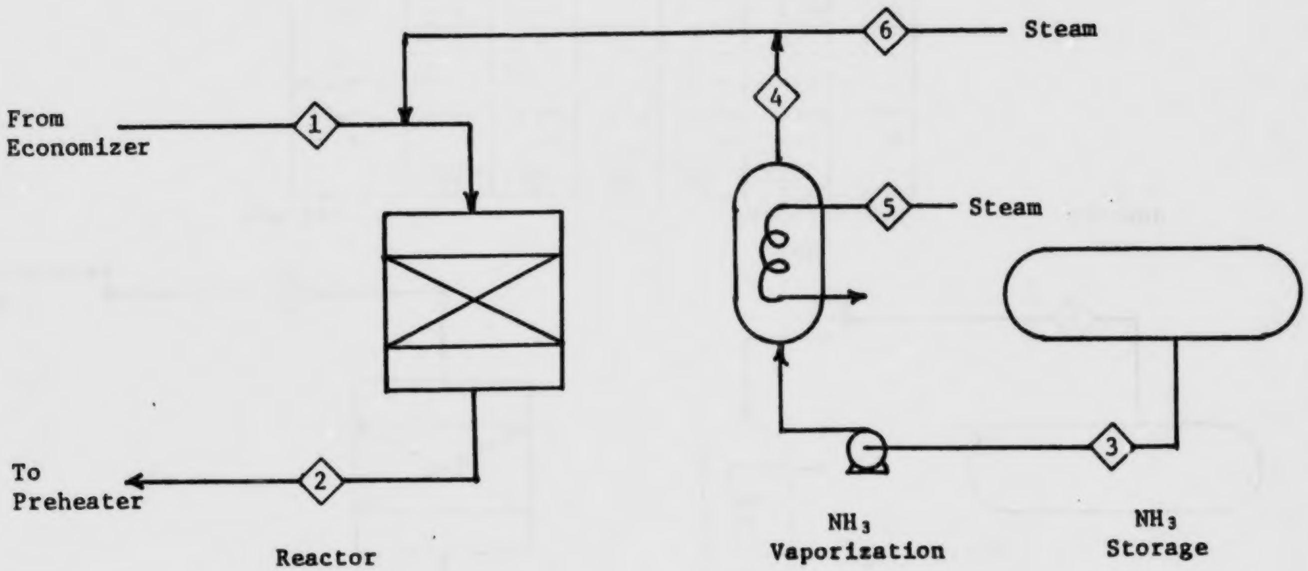
Natural Gas (4.4 MW)
Fixed Packed Bed SCR
Moderate Control


Steam
From Economizer
1
6
4
5
Steam
To Preheater
2
3
Reactor
NH₃ Vaporization
NH₃ Storage

| | ①| ② | ③ | ④ | ⑤ | ⑥ |
|---|---|---|---|---|---|---|
| T, °K | 450 | 450 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,850 | 615,000 | 752,000 | 552,000 | 310,000 |
| kg-mole/hr | | | | | | |
| $N_2$ | 183 | 183 | - | - | - | - |
| $CO_2$ | 21.4 | 21.4 | - | - | - | - |
| $H_2O$ | 46.6 | 46.7 | - | - | 0.010 | 0.088 |
| $O_2$ | 6.3 | 6.3 | - | - | - | - |
| $NO_x$ | 0.026 | 0.008 | - | - | - | - |
| $SO_x$ | tr | tr | - | - | - | - |
| $NH_3$ | - | <.001 | 0.018 | 0.018 | - | - |

MATERIAL BALANCE

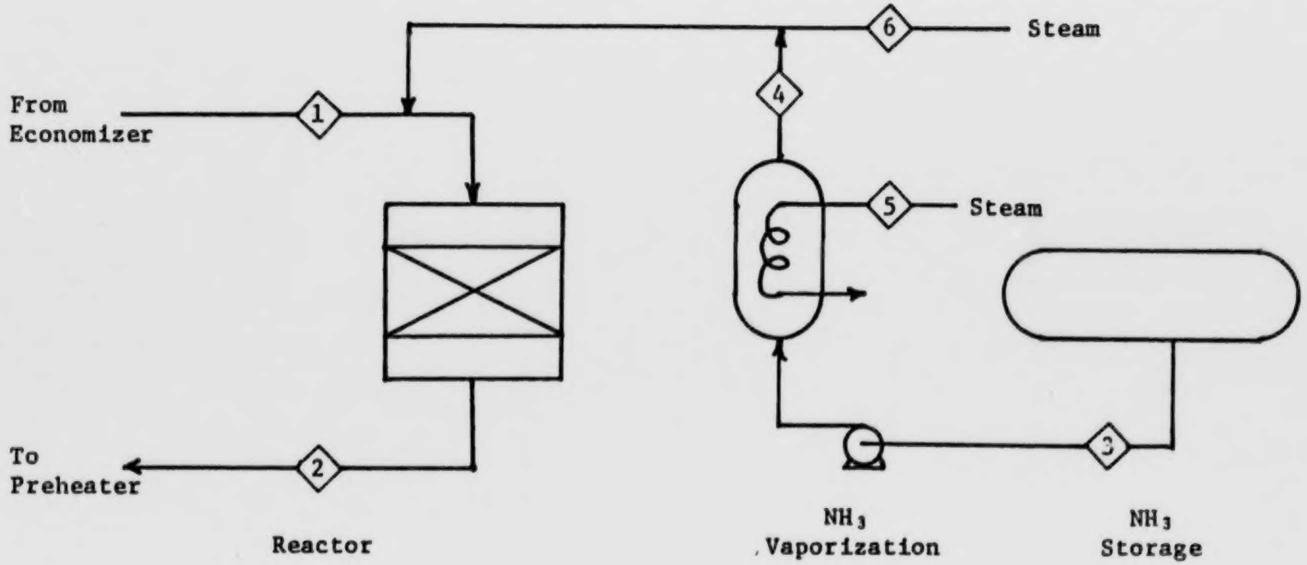
Natural Gas (44 MW)
Fixed Packed Bed SCR
Stringent Control


Steam
From
Economizer
1
6
4
Steam
5
To
Preheater
2
3
Reactor
$NH_3$
Vaporization
$NH_3$
Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K<br>P, Pa | | | | | | |
| $N_2$ | 1580 | 1580 | - | - | - | - |
| $CO_2$ | 185 | 185 | - | - | - | - |
| $H_2O$ | 402 | 404 | - | - | 0.146 | 1.32 |
| $O_2$ | 54.4 | 54.4 | - | - | - | - |
| $NO_x$ | 0.28 | 0.02 | - | - | - | - |
| $SO_x$ | tr | tr | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.233 | 0.233 | - | - |

## MATERIAL BALANCE

Natural Gas (44 MW)
Fixed Packed Bed SCR
Moderate Control


Steam
From Economizer
1
6
4
Steam
5
To Preheater
2
3
Reactor
$NH_3$ Vaporization
$NH_3$ Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K<br>P, Pa | | | | | | |
| $N_2$ | 1580 | 1580 | - | - | - | - |
| $CO_2$ | 185 | 185 | - | - | - | - |
| $H_2O$ | 402 | 404 | - | - | 0.113 | 1.02 |
| $O_2$ | 54.4 | 54.4 | - | - | - | - |
| $NO_x$ | 0.28 | 0.08 | - | - | - | - |
| $SO_x$ | tr | tr | - | - | - | - |
| $NH_3$ | - | <0.01 | 0.181 | 0.181 | - | - |

# APPENDIX 6

# CAPITAL COST BREAKDOWNS

## Table A6-1. CAPITAL COSTS

Boiler type: Underfeed
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Intermediate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 29,270 | | |
| Required auxiliaries | 24,200 | | |
| Total equipment cost | 53,470 | | |
| Installation costs, direct | | | |
| Foundations and supports | 4,290 | | |
| Piping | 14,080 | | |
| Insulation | 1,760 | | |
| Painting | 330 | | |
| Electrical | 1,350 | | |
| Instruments | 2,430 | | |
| Installation labor | 18,500 | | |
| Total installation cost | 42,740 | | |
| Total Direct Costs (equipment + installation) | | 96,210 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 9,621 | | |
| Construction fees | 9,621 | | |
| Start-up | 1,924 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 67,058 | |
| Contingencies | | 32,654 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 195,922 | |
| Land | | 490 | |
| Working capital | | 16,675 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $213,090 |

Table A6-2. CAPITAL COSTS

Boiler type: Chaingrate
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 61,510 | | |
| Required auxiliaries | 81,000 | | |
| Total equipment cost | 142,510 | | |
| Installation costs, direct | | | |
| Foundations and supports | 9,190 | | |
| Piping | 29,010 | | |
| Insulation | 3,900 | | |
| Painting | 700 | | |
| Electrical | 2,570 | | |
| Instruments | 4,820 | | |
| Installation labor | 41,310 | | |
| Total installation cost | 82,310 | | |
| Total Direct Costs (equipment + installation) | | 224,820 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 22,482 | | |
| Construction fees | 22,482 | | |
| Start-up | 4,496 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 95,352 | |
| Contingencies | | 64,034 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 384,206 | |
| Land | | 961 | |
| Working capital | | 28,129 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $413,300 |

Table A6-3. CAPITAL COSTS

Boiler type: Chaingrate
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Intermediate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 49,880 | | |
| Required auxiliaries | 60,600 | | |
| Total equipment cost | 110,480 | | |
| Installation costs, direct | | | |
| Foundations and supports | 7,430 | | |
| Piping | 23,470 | | |
| Insulation | 3,090 | | |
| Painting | 570 | | |
| Electrical | 2,100 | | |
| Instruments | 3,930 | | |
| Installation labor | 32,920 | | |
| Total installation cost | 73,510 | | |
| Total Direct Costs (equipment + installation) | | 183,990 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 18,399 | | |
| Construction fees | 18,399 | | |
| Start-up | 3,680 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 86,370 | |
| Contingencies | | 54,072 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 324,432 | |
| Land | | 811 | |
| Working capital | | 24,294 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $349,540 |

Table A6-4. CAPITAL COSTS

Boiler type: Chaingrate
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 42,000 | | |
| Required auxiliaries | 48,500 | | |
| Total equipment cost | 90,500 | | |
| Installation costs, direct | | | |
| Foundations and supports | 6,150 | | |
| Piping | 19,270 | | |
| Insulation | 2,490 | | |
| Painting | 480 | | |
| Electrical | 1,780 | | |
| Instruments | 3,260 | | |
| Installation labor | 27,170 | | |
| Total installation cost | 60,600 | | |
| Total Direct Costs (equipment + installation) | | 151,100 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 15,110 | | |
| Construction fees | 15,110 | | |
| Start-up | 3,022 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 79,134 | |
| Contingencies | | 46,047 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 276,281 | |
| Land | | 691 | |
| Working capital | | 21,894 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $298,870 |

Table A6-5. CAPITAL COSTS

Boiler type: Spreader Stoker
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Intermediate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 74,330 | | |
| Required auxiliaries | 121,000 | | |
| Total equipment cost | 195,330 | | |
| Installation costs, direct | | | |
| Foundations and supports | 10,260 | | |
| Piping | 32,610 | | |
| Insulation | 4,130 | | |
| Painting | 800 | | |
| Electrical | 2,950 | | |
| Instruments | 5,600 | | |
| Installation labor | 49,790 | | |
| Total installation cost | 106,140 | | |
| Total Direct Costs (equipment + installation) | | 301,470 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 30,147 | | |
| Construction fees | 30,147 | | |
| Start-up | 6,029 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 112,215 | |
| Contingencies | | 82,737 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 496,422 | |
| Land | | 1,241 | |
| Working capital | | 36,958 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $534,620 |

## Table A6-6. CAPITAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 104,470 | | |
| Required auxiliaries | 188,000 | | |
| Total equipment cost | 292,470 | | |
| Installation costs, direct | | | |
| Foundations and supports | 14,120 | | |
| Piping | 44,030 | | |
| Insulation | 5,560 | | |
| Painting | 1,090 | | |
| Electrical | 4,020 | | |
| Instruments | 7,770 | | |
| Installation labor | 69,860 | | |
| Total installation cost | 146,450 | | |
| Total Direct Costs (equipment + installation) | | 438,920 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 43,892 | | |
| Construction fees | 43,092 | | |
| Start-up | 8,778 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 142,454 | |
| Contingencies | | 116,275 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 697,649 | |
| Land | | 1,744 | |
| Working capital | | 53,013 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $752,410 |

## Table A6-7. CAPITAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 78,470 | | |
| Required auxiliaries | 113,000 | | |
| Total equipment cost | 191,470 | | |
| Installation costs, direct | | | |
| Foundations and supports | 10,540 | | |
| Piping | 33,160 | | |
| Insulation | 4,130 | | |
| Painting | 830 | | |
| Electrical | 3,050 | | |
| Instruments | 6,110 | | |
| Installation labor | 49,780 | | |
| Total installation cost | 107,600 | | |
| Total Direct Costs (equipment + installation) | | 299,070 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 29,907 | | |
| Construction fees | 29,907 | | |
| Start-up | 5,981 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 111,687 | |
| Contingencies | | 82,151 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 492,908 | |
| Land | | 1,232 | |
| Working capital | | 37,757 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $531,900 |

## Table A6-8. CAPITAL COSTS

Boiler type: Underfeed Stoker
Fuel: High Sulfur Eastern Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | |
|---|---|---|
| Total Direct Costs (equipment + installation) | | 1,284,000 |
| Installation costs, indirect | | |
| Engineering | 373,400 | |
| Construction and field expense | 128,400 | |
| Construction fees | 128,400 | |
| Start-up | 25,700 | |
| Performance tests | 4,000 | |
| Total Indirect Costs | | 660,000 |
| Contingencies | | 389,000 |
| Total Turnkey Costs (direct+indirect+contingencies) | | 2,333,000 |
| Land | | 6,000 |
| Working capital | | 64,000 |
| GRAND TOTAL (turnkey + land + working capital) | | $2,403,000 |

## Table A6-9. CAPITAL COSTS

Boiler type: Underfeed Stoker
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | | |
|---|---|---|---|
| Total Direct Costs (equipment + installation) | | 648,800 | |
| Installation costs, indirect | | | |
| Engineering | 373,400 | | |
| Construction and field expense | 64,900 | | |
| Construction fees | 64,900 | | |
| Start-up | 13,000 | | |
| Performance tests | 4,000 | | |
| Total Indirect Costs | | 520,200 | |
| Contingencies | | 233,800 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 1,403,000 | |
| Land | | 4,000 | |
| Working capital | | 43,000 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $1,450,000 |

## TABLE A6-10. CAPITAL COSTS

Boiler type: Pulverized Coal
Fuel: High Sulfur Eastern Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | |
|---|---|---|
| Total Direct Costs (equipment + installation) | | 3,734,000 |
| Installation costs, indirect | | |
| Engineering | 373,400 | |
| Construction and field expense | 373,400 | |
| Construction fees | 373,400 | |
| Start-up | 74,700 | |
| Performance tests | 4,000 | |
| Total Indirect Costs | | 1,199,000 |
| Contingencies | | 987,000 |
| Total Turnkey Costs (direct+indirect+contingencies) | | 5,920,000 |
| Land | | 15,000 |
| Working capital | | 180,000 |
| GRAND TOTAL (turnkey + land + working capital) | | $6,115,000 |

## TABLE A6-11. CAPITAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | |
|---|---|---|
| Total Direct Costs (equipment + installation) | | 1,793,000 |
| Installation costs, indirect | | |
| Engineering | 373,400 | |
| Construction and field expense | 179,300 | |
| Construction fees | 179,300 | |
| Start-up | 35,900 | |
| Performance tests | 4,000 | |
| Total Indirect Costs | | 772,000 |
| Contingencies | | 513,000 |
| Total Turnkey Costs (direct+indirect+contingencies) | | 3,078,000 |
| Land | | 8,000 |
| Working capital | | 78,000 |
| GRAND TOTAL (turnkey + land + working capital) | | $3,164,000 |

## Table A6-12. CAPITAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| **Equipment cost** | | | |
| Basic equipment (includes freight) | 8,230 | | |
| Required auxiliaries | 8,580 | | |
| Total equipment cost | 16,810 | | |
| **Installation costs, direct** | | | |
| Foundations and supports | 850 | | |
| Piping | 2,990 | | |
| Insulation | 360 | | |
| Painting | 60 | | |
| Electrical | 580 | | |
| Instruments | 470 | | |
| Installation labor | 3,910 | | |
| Total installation cost | 9,220 | | |
| Total Direct Costs (equipment + installation) | | 26,030 | |
| **Installation costs, indirect** | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 2,603 | | |
| Construction fees | 2,603 | | |
| Start-up | 521 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 51,619 | |
| Contingencies | | 11,647 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 89,296 | |
| Land | | 223 | |
| Working capital | | 9,892 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $99,410 |

Table A6-13. CAPITAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 6,230 | | |
| Required auxiliaries | 5,120 | | |
| Total equipment cost | 11,350 | | |
| Installation costs, direct | | | |
| Foundations and supports | 620 | | |
| Piping | 2,270 | | |
| Insulation | 280 | | |
| Painting | 40 | | |
| Electrical | 500 | | |
| Instruments | 350 | | |
| Installation labor | 2,980 | | |
| Total installation cost | 7,040 | | |
| Total Direct Costs (equipment + installation) | | 18,390 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 1,839 | | |
| Construction fees | 1,839 | | |
| Start-up | 368 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 49,938 | |
| Contingencies | | 10,248 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 78,576 | |
| Land | | 196 | |
| Working capital | | 9,402 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $88,170 |

## Table A6-14. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 34,030 | | |
| Required auxiliaries | 86,140 | | |
| Total equipment cost | 120,170 | | |
| Installation costs, direct | | | |
| Foundations and supports | 3,280 | | |
| Piping | 10,420 | | |
| Insulation | 1,150 | | |
| Painting | 220 | | |
| Electrical | 1,660 | | |
| Instruments | 1,690 | | |
| Installation labor | 21,300 | | |
| Total installation cost | 39,720 | | |
| Total Direct Costs (equipment + installation) | | 159,890 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 15,990 | | |
| Construction fees | 15,990 | | |
| Start-up | 3,200 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 81,070 | |
| Contingencies | | 36,140 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 277,100 | |
| Land | | 690 | |
| Working capital | | 27,810 | |
| GRAND TOTAL (turnkey + land + working capital) | | | 305,600 |

Table A6-15. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 25,400 | | |
| Required auxiliaries | 51,680 | | |
| Total equipment cost | 77,080 | | |
| Installation costs, direct | | | |
| Foundations and supports | 2,500 | | |
| Piping | 8,000 | | |
| Insulation | 900 | | |
| Painting | 180 | | |
| Electrical | 1,280 | | |
| Instruments | 1,320 | | |
| Installation labor | 14,930 | | |
| Total installation cost | 29,110 | | |
| Total Direct Costs (equipment + installation) | | 106,190 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 10,620 | | |
| Construction fees | 10,620 | | |
| Start-up | 2,120 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 69,250 | |
| Contingencies | | 26,320 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 201,760 | |
| Land | | 500 | |
| Working capital | | 21,510 | |
| GRAND TOTAL (turnkey + land + working capital) | | | 223,800 |

## Table A6-16. CAPITAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| **Equipment cost** | | | |
| Basic equipment (includes freight) | 29,100 | | |
| Required auxiliaries | 23,800 | | |
| Total equipment cost | 52,900 | | |
| **Installation costs, direct** | | | |
| Foundations and supports | 4,380 | | |
| Piping | 14,080 | | |
| Insulation | 1,870 | | |
| Painting | 460 | | |
| Electrical | 1,240 | | |
| Instruments | 2,350 | | |
| Installation labor | 18,580 | | |
| Total installation cost | 43,000 | | |
| Total Direct Costs (equipment + installation) | | 95,900 | |
| **Installation costs, indirect** | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 9,590 | | |
| Construction fees | 9,590 | | |
| Start-up | 1,920 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 67,000 | |
| Contingencies | | 24,400 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 187,300 | |
| Land | | 500 | |
| Working capital | | 15,100 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $202,900 |

Table A6-17. CAPITAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 20,910 | | |
| Required auxiliaries | 14,300 | | |
| Total equipment cost | 35,210 | | |
| Installation costs, direct | | | |
| Foundations and supports | 2,830 | | |
| Piping | 9,850 | | |
| Insulation | 1,300 | | |
| Painting | 240 | | |
| Electrical | 1,010 | | |
| Instruments | 1,670 | | |
| Installation labor | 12,820 | | |
| Total installation cost | 29,720 | | |
| Total Direct Costs (equipment + installation) | | 64,930 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 6,490 | | |
| Construction fees | 6,490 | | |
| Start-up | 1,300 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 60,170 | |
| Contingencies | | 18,770 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 143,900 | |
| Land | | 400 | |
| Working capital | | 13,800 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $158,100 |

Table A6-18. CAPITAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 21,340 | | |
| Required auxiliaries | 15,740 | | |
| Total equipment cost | 37,080 | | |
| Installation costs, direct | | | |
| Foundations and supports | 3,070 | | |
| Piping | 9,870 | | |
| Insulation | 1,290 | | |
| Painting | 250 | | |
| Electrical | 1,010 | | |
| Instruments | 1,710 | | |
| Installation labor | 12,100 | | |
| Total installation cost | 29,300 | | |
| Total Direct Costs (equipment + installation) | | 66,380 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 6,640 | | |
| Construction fees | 6,640 | | |
| Start-up | 1,330 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 60,500 | |
| Contingencies | | 19,030 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 145,910 | |
| Land | | 400 | |
| Working capital | | 20,350 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $166,700 |

## Table A6-19. CAPITAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 17,440 | | |
| Required auxiliaries | 9,440 | | |
| Total equipment cost | 26,880 | | |
| Installation costs, direct | | | |
| Foundations and supports | 2,540 | | |
| Piping | 8,090 | | |
| Insulation | 1,060 | | |
| Painting | 200 | | |
| Electrical | 870 | | |
| Instruments | 1,390 | | |
| Installation labor | 9,810 | | |
| Total installation cost | 23,960 | | |
| Total Direct Costs (equipment + installation) | | 50,840 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 5,080 | | |
| Construction fees | 5,080 | | |
| Start-up | 1,020 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 57,070 | |
| Contingencies | | 16,190 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 124,100 | |
| Land | | 300 | |
| Working capital | | 19,190 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $144,600 |

## Table A6-20. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 74,670 | | |
| Required auxiliaries | 113,000 | | |
| Total equipment cost | 187,670 | | |
| Installation costs, direct | | | |
| Foundations and supports | 10,770 | | |
| Piping | 34,420 | | |
| Insulation | 4,620 | | |
| Painting | 820 | | |
| Electrical | 3,180 | | |
| Instruments | 5,660 | | |
| Installation labor | 50,720 | | |
| Total installation cost | 110,190 | | |
| Total Direct Costs (equipment + installation) | | 297,860 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 29,786 | | |
| Construction fees | 29,786 | | |
| Start-up | 5,957 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 111,421 | |
| Contingencies | | 61,392 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 470,673 | |
| Land | | 1,177 | |
| Working capital | | 31,290 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $503,140 |

## Table A6-21. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 54,980 | | |
| Required auxiliaries | 68,100 | | |
| Total equipment cost | 123,080 | | |
| Installation costs, direct | | | |
| Foundations and supports | 7,900 | | |
| Piping | 25,320 | | |
| Insulation | 3,310 | | |
| Painting | 600 | | |
| Electrical | 2,360 | | |
| Instruments | 4,310 | | |
| Installation labor | 35,890 | | |
| Total installation cost | 79,690 | | |
| Total Direct Costs (equipment + installation) | | 202,770 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 20,277 | | |
| Construction fees | 20,277 | | |
| Start-up | 4,055 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 90,501 | |
| Contingencies | | 43,991 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 337,262 | |
| Land | | 843 | |
| Working capital | | 23,456 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $378,190 |

## Table A6-22. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 34,650 | | |
| Required auxiliaries | 89,550 | | |
| Total equipment cost | 124,200 | | |
| Installation costs, direct | | | |
| Foundations and supports | 6,620 | | |
| Piping | 20,870 | | |
| Insulation | 2,650 | | |
| Painting | 510 | | |
| Electrical | 2,050 | | |
| Instruments | 3,620 | | |
| Installation labor | 26,110 | | |
| Total installation cost | 62,430 | | |
| Total Direct Costs (equipment + installation) | | 186,630 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 18,663 | | |
| Construction fees | 18,663 | | |
| Start-up | 3,733 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 86,951 | |
| Contingencies | | 41,037 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 314,617 | |
| Land | | 787 | |
| Working capital | | 31,621 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $347,020 |

Table A6-23. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 26,170 | | |
| Required auxiliaries | 57,550 | | |
| Total equipment cost | 83,720 | | |
| Installation costs, direct | | | |
| Foundations and supports | 4,740 | | |
| Piping | 14,930 | | |
| Insulation | 1,880 | | |
| Painting | 370 | | |
| Electrical | 1,510 | | |
| Instruments | 2,650 | | |
| Installation labor | 18,570 | | |
| Total installation cost | 44,650 | | |
| Total Direct Costs (equipment + installation) | | 128,370 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 12,837 | | |
| Construction fees | 12,837 | | |
| Start-up | 2,567 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 74,133 | |
| Contingencies | | 30,376 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 232,883 | |
| Land | | 582 | |
| Working capital | | 26,223 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $259,690 |

Table A6-24. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR ($NO_x$/$SO_x$)
Control level: Intermediate

| | | |
|---|---|---|
| Total Direct Costs (equipment + installation) | | 2,323,000 |
| Installation costs, indirect | | |
| Engineering | 373,400 | |
| Construction and field expense | 232,300 | |
| Construction fees | 232,300 | |
| Start-up | 46,500 | |
| Performance tests | 4,000 | |
| Total Indirect Costs | | 889,000 |
| Contingencies | | 482,000 |
| Total Turnkey Costs (direct+indirect+contingencies) | | 3,693,000 |
| Land | | 9,000 |
| Working capital | | 99,000 |
| GRAND TOTAL (turnkey + land + working capital) | | $3,801,000 |

## Table A6-25. CAPITAL COSTS

Boiler type: Firetube (4.4 $\text{MW}_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 8,530 | | |
| Required auxiliaries | 8,900 | | |
| Total equipment cost | 17,430 | | |
| Installation costs, direct | | | |
| Foundations and supports | 870 | | |
| Piping | 3,050 | | |
| Insulation | 360 | | |
| Painting | 60 | | |
| Electrical | 590 | | |
| Instruments | 470 | | |
| Installation labor | 3,990 | | |
| Total installation cost | 9,390 | | |
| Total Direct Costs (equipment + installation) | | 26,820 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 2,682 | | |
| Construction fees | 2,682 | | |
| Start-up | 536 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 51,792 | |
| Contingencies | | 11,792 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 90,404 | |
| Land | | 226 | |
| Working capital | | 9,893 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $100,520 |

## Table A6-26. CAPITAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 6,580 | | |
| Required auxiliaries | 5,320 | | |
| Total equipment cost | 11,900 | | |
| Installation costs, direct | | | |
| Foundations and supports | 650 | | |
| Piping | 2,330 | | |
| Insulation | 280 | | |
| Painting | 40 | | |
| Electrical | 510 | | |
| Instruments | 360 | | |
| Installation labor | 5,590 | | |
| Total installation cost | 9,760 | | |
| Total Direct Costs (equipment + installation) | | 21,660 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 2,166 | | |
| Construction fees | 2,166 | | |
| Start-up | 433 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 50,657 | |
| Contingencies | | 10,848 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 83,165 | |
| Land | | 208 | |
| Working capital | | 9,391 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $92,760 |

## Table A6-27. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 33,410 | | |
| Required auxiliaries | 87,700 | | |
| Total equipment cost | 121,110 | | |
| Installation costs, direct | | | |
| Foundations and supports | 3,310 | | |
| Piping | 10,480 | | |
| Insulation | 1,160 | | |
| Painting | 220 | | |
| Electrical | 1,670 | | |
| Instruments | 1,710 | | |
| Installation labor | 21,380 | | |
| Total installation cost | 39,930 | | |
| Total Direct Costs (equipment + installation) | | 161,040 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 16,100 | | |
| Construction fees | 16,100 | | |
| Start-up | 3,220 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 81,310 | |
| Contingencies | | 36,350 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 278,700 | |
| Land | | 700 | |
| Working capital | | 27,440 | |
| GRAND TOTAL (turnkey + land + working capital) | | | 306,800 |

## Table A6-28. CAPITAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 25,760 | | |
| Required auxiliaries | 52,580 | | |
| Total equipment cost | 78,340 | | |
| Installation costs, direct | | | |
| Foundations and supports | 2,520 | | |
| Piping | 8,050 | | |
| Insulation | 900 | | |
| Painting | 180 | | |
| Electrical | 1,280 | | |
| Instruments | 1,340 | | |
| Installation labor | 15,000 | | |
| Total installation cost | 29,270 | | |
| Total Direct Costs (equipment + installation) | | 107,610 | |
| Installation costs, indirect | | | |
| Engineering | 43,890 | | |
| Construction and field expense | 10,760 | | |
| Construction fees | 10,760 | | |
| Start-up | 2,150 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 69,560 | |
| Contingencies | | 26,580 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 203,750 | |
| Land | | 510 | |
| Working capital | | 19,470 | |
| GRAND TOTAL (turnkey + land + working capital) | | | 223,700 |

# APPENDIX 7

# ANNUAL COST BREAKDOWNS

## Table A7-1. ANNUAL COSTS

Boiler type: Underfeed
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 3,527 | | |
| Catalyst | 14,501 | | |
| Electricity | 854 | | |
| Steam | 776 | | |
| Ammonia | 2,431 | | |
| Total direct cost | | 66,699 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 12,516 | | |
| Total overhead cost | | 17,250 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 7,837 | | |
| Capital recovery | 25,750 | | |
| Total capital charges | | 33,595 | |
| TOTAL ANNUALIZED COSTS | | | $117,540 |

Table A7-2. ANNUAL COSTS

Boiler type: Chaingrate
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 6,916 | | |
| Catalyst | 48,590 | | |
| Electricity | 3,101 | | |
| Steam | 2,272 | | |
| Ammonia | 7,030 | | |
| Total direct cost | | 112,519 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 13,396 | | |
| Total overhead cost | | 18,130 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 15,368 | | |
| Capital recovery | 50,512 | | |
| Total capital charges | | 65,880 | |
| TOTAL ANNUALIZED COSTS | | | $196,530 |

## Table A7-3. ANNUAL COSTS

Boiler type: Chaingrate
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 5,840 | | |
| Catalyst | 36,329 | | |
| Electricity | 2,345 | | |
| Steam | 1,943 | | |
| Ammonia | 6,110 | | |
| Total direct cost | | 97,177 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 13,117 | | |
| Total overhead cost | | 17,851 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 12,977 | | |
| Capital recovery | 42,653 | | |
| Total capital charges | | 55,630 | |
| TOTAL ANNUALIZED COSTS | | | $170,660 |

Table A7-4. ANNUAL COSTS

Boiler type: Chaingrate
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 4,973 | | |
| Catalyst | 29,129 | | |
| Electricity | 1,892 | | |
| Steam | 1,713 | | |
| Ammonia | 5,256 | | |
| Total direct cost | | 87,573 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 12,892 | | |
| Total overhead cost | | 17,626 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 11,051 | | |
| Capital recovery | 36,323 | | |
| Total capital charges | | 47,374 | |
| TOTAL ANNUALIZED COSTS | | | $152,570 |

## Table A7-5. ANNUAL COSTS

Boiler type: Spreader Stoker
Fuel: Low Sulfur Western Coal
Control technique: Farallel Flow SCR
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 8,936 | | |
| Catalyst | 72,600 | | |
| Electricity | 5,847 | | |
| Steam | 3,814 | | |
| Ammonia | 12,023 | | |
| Total direct cost | | 147,830 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 13,922 | | |
| Total overhead cost | | 18,656 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 19,857 | | |
| Capital recovery | 65,265 | | |
| Total capital charges | | 85,122 | |
| TOTAL ANNUALIZED COSTS | | | $251,610 |

## Table A7-6. ANNUAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $ 15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 12,558 | | |
| Catalyst | 112,800 | | |
| Electricity | 12,417 | | |
| Steam | 7,133 | | |
| Ammonia | 22,535 | | |
| Total direct cost | | 212,053 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 14,864 | | |
| Total overhead cost | | 19,598 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 27,906 | | |
| Capital recovery | 91,720 | | |
| Total capital charges | | 119,626 | |
| TOTAL ANNUALIZED COSTS | | | $351,280 |

Table A7-7. ANNUAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $15,780 | | |
| Maintenance labor | 28,830 | | |
| Materials | 8,872 | | |
| Catalyst | 67,800 | | |
| Electricity | 7,490 | | |
| Steam | 5,371 | | |
| Ammonia | 16,885 | | |
| Total direct cost | | 151,028 | |
| Overhead | | | |
| Payroll | 4,734 | | |
| Plant | 13,905 | | |
| Total overhead cost | | 18,639 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 19,716 | | |
| Capital recovery | 64,803 | | |
| Total capital charges | | 84,519 | |
| TOTAL ANNUALIZED COSTS | | | $254,190 |

Table A7-8. ANNUAL COSTS

Boiler type: Underfeed Stoker
Fuel: High Sulfur Eastern Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 31,590 | | |
| Maintenance labor | 76,900 | | |
| Materials | 70,000 | | |
| Catalyst | 12,450 | | |
| Electricity | 19,730 | | |
| Steam | 26,100 | | |
| Fuel | 33,790 | | |
| Boiler feed water | 39,520 | | |
| Ammonia | 3,030 | | |
| Heat credit | (-32,270) | | |
| By-product credit | (-26,050) | | |
| Total direct cost | | 254,800 | |
| Overhead | | | |
| Payroll | 9,480 | | |
| Plant | 46,410 | | |
| Total overhead cost | | 55,900 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 93,320 | | |
| Capital recovery | 306,700 | | |
| Total capital charges | | 400,000 | |
| TOTAL ANNUALIZED COSTS | | | $710,700 |

## Table A7-9. ANNUAL COSTS

Boiler type: Underfeed Stoker
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 31,590 | | |
| Maintenance labor | 76,900 | | |
| Materials | 42,090 | | |
| Catalyst | 2,620 | | |
| Electricity | 6,990 | | |
| Steam | 6,400 | | |
| Fuel | 7,140 | | |
| Boiler feed water | 8,350 | | |
| Ammonia | 3,710 | | |
| Heat credit | (-7,780) | | |
| By-product credit | (-5,490) | | |
| Total direct cost | | 172,500 | |
| Overhead | | | |
| Payroll | 9,480 | | |
| Plant | 39,150 | | |
| Total overhead cost | | 48,600 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 56,120 | | |
| Capital recovery | 184,450 | | |
| Total capital charges | | 240,600 | |
| TOTAL ANNUALIZED COSTS | | | $462,000 |

Table A7-10. ANNUAL COSTS

Boiler type: Pulverized Coal
Fuel: High Sulfur Eastern Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 31,590 | | |
| Maintenance labor | 76,900 | | |
| Materials | 106,600 | | |
| Catalyst | 83,100 | | |
| Electricity | 128,100 | | |
| Steam | 175,200 | | |
| Fuel | 226,500 | | |
| Boiler feed water | 269,500 | | |
| Ammonia | 24,100 | | |
| Heat credit | (-223,000) | | |
| By-product credit | (-173,900) | | |
| Total direct cost | | 724,700 | |
| Overhead | | | |
| Payroll | 9,480 | | |
| Plant | 55,920 | | |
| Total overhead cost | | 65,400 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 236,800 | | |
| Capital recovery | 778,300 | | |
| Total capital charges | | 1,010,000 | |
| TOTAL ANNUALIZED COSTS | | | $1,805,000 |

## Table A7-11. ANNUAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR ($NO_x/SO_x$)
Control level: Intermediate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 31,590 | | |
| Maintenance labor | 76,900 | | |
| Materials | 55,400 | | |
| Catalyst | 17,520 | | |
| Electricity | 43,560 | | |
| Steam | 44,140 | | |
| Fuel | 47,620 | | |
| Boiler feed water | 55,820 | | |
| Ammonia | 29,780 | | |
| Heat credit | (-51,930) | | |
| By-product credit | (-36,680) | | |
| Total direct cost | | 313,700 | |
| Overhead | | | |
| Payroll | 9,480 | | |
| Plant | 42,610 | | |
| Total overhead cost | | 52,100 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 123,100 | | |
| Capital recovery | 404,700 | | |
| Total capital charges | | 527,800 | |
| TOTAL ANNUALIZED COSTS | | | $893,600 |

## Table A7-12. ANNUAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 11,835 | | |
| Maintenance labor | 21,623 | | |
| Materials | 1,205 | | |
| Catalyst | 3,861 | | |
| Electricity | 547 | | |
| Steam | 301 | | |
| Ammonia | 197 | | |
| Total direct cost | | 39,569 | |
| Overhead | | | |
| Payroll | 3,551 | | |
| Plant | 9,012 | | |
| Total overhead cost | | 12,563 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 3,571 | | |
| Capital recovery | 11,740 | | |
| Total capital charges | | 15,311 | |
| TOTAL ANNUALIZED COSTS | | | $67,440 |

Table A7-13. ANNUAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $11,835 | | |
| Maintenance labor | 21,623 | | |
| Materials | 1,061 | | |
| Catalyst | 2,304 | | |
| Electricity | 345 | | |
| Steam | 291 | | |
| Ammonia | 148 | | |
| Total direct cost | | 37,607 | |
| Overhead | | | |
| Payroll | 3,551 | | |
| Plant | 8,975 | | |
| Total overhead cost | | 12,526 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 3,143 | | |
| Capital recovery | 10,331 | | |
| Total capital charges | | 13,474 | |
| TOTAL ANNUALIZED COSTS | | | $63,610 |

## Table A7-14. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,480 | | |
| Maintainence labor | 26,440 | | |
| Materials | 8,310 | | |
| Catalyst | 47,380 | | |
| Electricity | 8,500 | | |
| Steam | 3,740 | | |
| Ammonia | 2,470 | | |
| Total direct cost | | 111,270 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 12,800 | | |
| Total overhead cost | | 17,140 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 11,080 | | |
| Capital recovery | 36,430 | | |
| Total capital charges | | 47,510 | |
| TOTAL ANNUALIZED COSTS | | | 175,900 |

### Table A7-15. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Distillate Oil
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,480 | | |
| Maintainence labor | 26,440 | | |
| Materials | 6,050 | | |
| Catalyst | 28,420 | | |
| Electricity | 5,140 | | |
| Steam | 3,570 | | |
| Ammonia | 1,920 | | |
| Total direct cost | | 86,020 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 12,210 | | |
| Total overhead cost | | 16,550 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 8,070 | | |
| Capital recovery | 26,530 | | |
| Total capital charges | | 34,600 | |
| TOTAL ANNUALIZED COSTS | | | 137,200 |

Table A7-16. ANNUAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,480 | | |
| Maintenance labor | 26,430 | | |
| Materials | 3,090 | | |
| Catalyst | 13,090 | | |
| Electricity | 740 | | |
| Steam | 590 | | |
| Ammonia | 1,840 | | |
| Total direct cost | | 60,260 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 11,440 | | |
| Total overhead cost | | 15,780 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 7,490 | | |
| Capital recovery | 24,620 | | |
| Total capital charges | | 32,110 | |
| TOTAL ANNUALIZED COSTS | | | $108,200 |

## Table A7-17. ANNUAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,480 | | |
| Maintenance labor | 26,430 | | |
| Materials | 4,320 | | |
| Catalyst | 7,870 | | |
| Electricity | 460 | | |
| Steam | 440 | | |
| Ammonia | 1,370 | | |
| Total direct cost | | 55,370 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 11,760 | | |
| Total overhead cost | | 16,100 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 5,750 | | |
| Capital recovery | 18,910 | | |
| Total capital charges | | 24,660 | |
| TOTAL ANNUALIZED COSTS | | | $96,100 |

Table A7-18. ANNUAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,465 | | |
| Maintenance labor | 52,855 | | |
| Materials | 2,410 | | |
| Catalyst | 8,660 | | |
| Electricity | 570 | | |
| Steam | 590 | | |
| Ammonia | 1,840 | | |
| Total direct cost | | 81,390 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 18,130 | | |
| Total overhead cost | | 22,470 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 5,840 | | |
| Capital recovery | 19,180 | | |
| Total capital charges | | 25,020 | |
| TOTAL ANNUALIZED COSTS | | | $129,900 |

## Table A7-19. ANNUAL COSTS

Boiler type: Watertube (8.8 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,465 | | |
| Maintenance labor | 52,855 | | |
| Materials | 2,050 | | |
| Catalyst | 5,190 | | |
| Electricity | 400 | | |
| Steam | 440 | | |
| Ammonia | 1,370 | | |
| Total direct cost | | 76,770 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 18,040 | | |
| Total overhead cost | | 22,380 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 4,960 | | |
| Capital recovery | 16,320 | | |
| Total capital charges | | 21,280 | |
| TOTAL ANNUALIZED COSTS | | | $120,400 |

Table A7-20. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $14,465 | | |
| Maintenance labor | 26,428 | | |
| Materials | 7,766 | | |
| Catalyst | 62,150 | | |
| Electricity | 5,697 | | |
| Steam | 2,089 | | |
| Ammonia | 6,565 | | |
| Total direct cost | | 125,160 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 12,651 | | |
| Total overhead cost | | 16,991 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 18,827 | | |
| Capital recovery | 61,879 | | |
| Total capital charges | | 80,706 | |
| TOTAL ANNUALIZED COSTS | | | $222,860 |

## Table A7-21. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $14,465 | | |
| Maintenance labor | 26,428 | | |
| Materials | 5,565 | | |
| Catalyst | 37,455 | | |
| Electricity | 3,457 | | |
| Steam | 1,516 | | |
| Ammonia | 4,938 | | |
| Total direct cost | | 93,824 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 12,079 | | |
| Total overhead cost | | 16,419 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 13,490 | | |
| Capital recovery | 44,340 | | |
| Total capital charges | | 57,830 | |
| TOTAL ANNUALIZED COSTS | | | $181,180 |

## Table A7-22. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $14,465 | | |
| Maintenance labor | 52,855 | | |
| Materials | 5,191 | | |
| Catalyst | 41,510 | | |
| Electricity | 3,820 | | |
| Steam | 2,080 | | |
| Ammonia | 6,560 | | |
| Total direct cost | | 126,481 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 18,853 | | |
| Total overhead cost | | 23,193 | |
| Capital Charges | | | |
| C&A, taxes & ins. | 12,585 | | |
| Capital recovery | 41,363 | | |
| Total capital charges | | 53,948 | |
| TOTAL ANNUALIZED COSTS | | | $203,620 |

### Table A7-23. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Moving Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $14,465 | | |
| Maintenance labor | 52,855 | | |
| Materials | 3,842 | | |
| Catalyst | 24,900 | | |
| Electricity | 2,350 | | |
| Steam | 1,540 | | |
| Ammonia | 4,940 | | |
| Total direct cost | | 104,892 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 18,502 | | |
| Total overhead cost | | 22,842 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 9,315 | | |
| Capital recovery | 30,617 | | |
| Total capital charges | | 39,932 | |
| TOTAL ANNUALIZED COSTS | | | $167,670 |

Table A7-24. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Residual Oil
Control technique: Parallel Flow SCR ($NO_x$/$SO_x$)
Control level: Intermediate

| | | |
|---|---|---|
| Direct costs | | |
| Direct labor | 28,960 | |
| Maintenance labor | 70,490 | |
| Materials | 110,790 | |
| Catalyst | 31,900 | |
| Electricity | 53,330 | |
| Steam | 66,780 | |
| Fuel | 86,490 | |
| Boiler feed water | 102,980 | |
| Ammonia | 8,710 | |
| Heat credit | (-86,530) | |
| By-product credit | (-78,290) | |
| Total direct cost | | 395,600 |
| Overhead | | |
| Payroll | 8,690 | |
| Plant | 54,660 | |
| Total overhead cost | | 63,350 |
| Capital Charges | | |
| G&A, taxes & ins. | 147,720 | |
| Capital recovery | 485,520 | |
| Total capital charges | | 633,240 |
| TOTAL ANNUALIZED COSTS | | $1,092,000 |

## Table A7-25. ANNUAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $11,835 | | |
| Maintenance labor | 21,623 | | |
| Materials | 1,220 | | |
| Catalyst | 4,005 | | |
| Electricity | 600 | | |
| Steam | 70 | | |
| Ammonia | 220 | | |
| Total direct cost | | 39,573 | |
| Overhead | | | |
| Payroll | 3,551 | | |
| Plant | 9,016 | | |
| Total overhead cost | | 12,567 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 3,616 | | |
| Capital recovery | 11,885 | | |
| Total capital charges | | 15,501 | |
| TOTAL ANNUALIZED COSTS | | | $67,640 |

Table A7-26. ANNUAL COSTS

Boiler type: Firetube (4.4 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | $11,835 | | |
| Maintenance labor | 21,623 | | |
| Materials | 1,123 | | |
| Catalyst | 2,394 | | |
| Electricity | 370 | | |
| Steam | 50 | | |
| Ammonia | 170 | | |
| Total direct cost | | 37,565 | |
| Overhead | | | |
| Payroll | 3,551 | | |
| Plant | 8,991 | | |
| Total overhead cost | | 12,542 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 3,327 | | |
| Capital recovery | 19,934 | | |
| Total capital charges | | 14,261 | |
| TOTAL ANNUALIZED COSTS | | | $64,370 |

## Table A7-27. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,480 | | |
| Maintainence labor | 26,440 | | |
| Materials | 8,360 | | |
| Catalyst | 48,240 | | |
| Electricity | 8,630 | | |
| Steam | 880 | | |
| Ammonia | 2,730 | | |
| Total direct cost | | 109,760 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 12,810 | | |
| Total overhead cost | | 17,510 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 11,150 | | |
| Capital recovery | 36,640 | | |
| Total capital charges | | 47,790 | |
| TOTAL ANNUALIZED COSTS | | | 174,700 |

## Table A7-28. ANNUAL COSTS

Boiler type: Watertube (44 $MW_t$)
Fuel: Natural Gas
Control technique: Fixed Packed Bed SCR
Control level: Moderate

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 14,480 | | |
| Maintainence labor | 26,440 | | |
| Materials | 6,110 | | |
| Catalyst | 28,920 | | |
| Electricity | 5,230 | | |
| Steam | 680 | | |
| Ammonia | 2,120 | | |
| Total direct cost | | 77,870 | |
| Overhead | | | |
| Payroll | 4,340 | | |
| Plant | 12,230 | | |
| Total overhead cost | | 16,570 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 8,150 | | |
| Capital recovery | 26,790 | | |
| Total capital charges | | 34,940 | |
| TOTAL ANNUALIZED COSTS | | | 129,400 |

# APPENDIX 8

# SAMPLE CALCULATIONS

An example calculation is shown below to illustrate how the energy vaues were arrived at. The example illustrates the case of a Pulverized Coal standard boiler with a Parallel Flow reactor and stringent control. Calculations for the other standard boilers were performed in a similar manner.

Sample Calculation--

First, it is necessary to perform a combustion calculation to characterize the flue gas.

130% excess air

Basis: Coal analysis: lb/lb fuel fired

C = 0.5760
$H_2$ = 0.320
$O_2$ = 0.1120
$N_2$ = 0.0120
S = 0.0060
$H_2O$ = 0.2080
Ash = 0.0540
1.0000

The calculation is based on a method presented in Steam[1] and the values shown here are documented in the reference.

## $O_2$ and Air required for combustion

| | | $O_2$, $\frac{\text{lb}}{\text{lb fuel fired}}$ | Air, $\frac{\text{lb}}{\text{lb fuel fired}}$ |
|---|---|---|---|
| C | 0.5760 | x 2.66 = 1.532 , | x 11.53 = 6.641 |
| $H_2$ | 0.0320 | x 7.94 = 0.254 , | x 34.34 = 1.099 |
| $O_2$ | 0.1120 | | |
| $N_2$ | 0.120 | | |
| S | 0.0060 | x 1.00 = 0.006 , | x 4.29 = 0.026 |
| $H_2O$ | 0.2080 | | |
| Ash | 0.0540 | | |
| Total | 1.00 | 1.792 | 7.766 |
| Less $O_2$ in the fuel | | − 0.112 | − 0.482 |
| Requirement | | 1.680 | 7.284 |
| Excess (30%) | | 0.504 | 2.185 |
| Total | | 2.184 | 9.469 |

## Products of combustion

| | | | lb/lb fuel |
|---|---|---|---|
| $CO_2$ | 0.5760 x 3.66 | = | 2.108 |
| $H_2O$ | (0.032 x 8.94) + (0.2080) + (0.013 x 9.469) | = | 0.617 |
| $O_2$ | excess | = | 0.504 |
| $N_2$ | 9.469 x (0.7685 + 0.0120) | = | 7.391 |
| $NO_x$ | (specified by Acurex) | = | 0.0090 |
| $SO_x$ | (specified by Acurex) | = | 0.0114 |
| CO | (specified by Acurex) | = | 0.0005 |
| HC as $CH_4$ | (specified by Acurex) | = | 0.0001 |
| Fly ash | (specified by Acurex) | = | 0.0432 |
| | | | 10.68 |

$$\text{Fuel feed rate} = 10.42\ \frac{\text{tons}}{\text{hr}} \times \frac{2000\ \text{lb}}{\text{ton}} = 20{,}840\ \frac{\text{lb fuel}}{\text{hr}}$$

Flue Gas Composition:

| | lb/hr | moles/hr | mole % |
|---|---|---|---|
| $N_2$ | 154,000 | 5,496 | 72.8 |
| $CO_2$ | 43,900 | 998 | 13.2 |
| $H_2O$ | 12,900 | 716 | 9.5 |
| $O_2$ | 10,500 | 328 | 4.3 |
| $SO_2$ | 238 | 4 | 0.1 |
| $NO_x$ | 188 | 4 | 0.1 |
| CO | 10 | - | - |
| HC (as $CH_4$) | 3 | - | - |
| Fly ash | 900 | - | - |
| | 222,639 | 7,546 | 100.0 |

Average molecular weight $= \frac{222,639}{7,546} = 29.5 \frac{\text{lb}}{\text{lb mole}}$

$$= 29.5 \frac{\text{g}}{\text{g mole}}$$

Flue gas flow rate

$$G_v = \left(73,200 \frac{\text{ft}^3}{\text{min}}\right)\left(\frac{492°\text{R}}{810°\text{R}}\right)\left(\frac{1.698\ \text{Nm}^3/\text{hr}}{\text{scfm}}\right) = 75,500 \frac{\text{Nm}^3}{\text{hr}}$$

### Reactor Sizing

Next, it is necessary to size the reactor so the pressure drop across the reactor may be calculated. For the stringent level of control, a large reactor size and bed depth are used to ensure 90% $NO_x$ reduction.

Basis:[2] Space velocity = 3000 $\text{hr}^{-1}$ (based on catalyst volume: 3000 catalyst volumes of flue gas per hour)

Bed depth = 4.5 m

Catalyst volume$= \frac{75,500\ \text{Nm}^3/\text{hr}}{3,000\ \text{hr}^{-1}} = 25.2\ \text{m}^3$

To calculate the reactor volume, the specific surface areas of the pure catalyst and of the catalyst packed in a reactor are needed.

601 $m^2/m^3$ catalyst
194 $m^2/m^3$ packed reactor volume
(20mm parallel plate)

$$\text{Reactor volume} = (25.2\ m^3\ \text{catalyst})\left(\frac{601\ m^2/m^3\ \text{catalyst volume}}{194\ m^2/m^3\ \text{packed reactor volume}}\right)$$

$$= 78\ m^3$$

$$= \text{width}^2 \times \text{depth (square reactor)}$$

Therefore, width = 4.16 m

## Pressure Drop

Now that the reactor geometry has been defined, the pressure drop across the reactor can be determined. For this calculation the following equation is used.[3]

$$\Delta P = \frac{2\ f_m G^2 L\ (1-\varepsilon)^{3-n}}{D_p g_c \rho \phi_s^{3-n} \varepsilon^3} \qquad \text{(A8-1)}$$

where $\Delta P$ = pressure drop across bed of granular solids, $lb_f/ft^2$

$f_m$ = friction factor: a function of modified Reynold's number ($N'_{Re}$), dimensionless

$G$ = gas superficial mass velocity, $lb_m/ft^2$ sec

$L$ = bed depth, ft

$\varepsilon$ = void fraction, dimensionless

$n$ = exponent: a function of modified Reynold's number ($N'_{Re}$), dimensionless

$D_p$ = average particle diameter: diameter of a sphere of the same volume as the non-spherical particle, ft

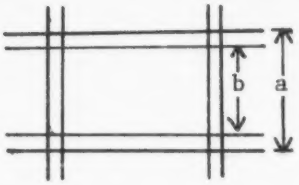
$g_c$ = dimensional constant, 32.2 $\frac{lb_m\ ft}{sec^2 lb_f}$

$\rho$ = gas density, $lb_m/ft^3$

$\phi_s$ = shape factor of the solid: quotient of the surface area of a sphere of equivalent volume divided by the actual surface area of the non-spherical particle, dimensionless

(The modified Reynolds number, $N'_{Re}$, is defined as $D_pG/\mu$, where $\mu$ = gas viscosity, $lb_m/ft$ sec)

Parallel Flow Catalyst (a square passage was assumed for ease of calculation)



a = 20 mm

b = 14 mm

Cell length = 1 m (assumed, a common commercial cell length[2])

In order to calculate a shape factor it is necessary to calculate the diameter of a sphere that has a volume of catalyst equivalent to a single passage of the square honeycomb catalyst.

$$\text{Catalyst volume per passage} = \left[(20\ \text{mm})^2 - (14\ \text{mm})^2\right] 1000\ \text{mm}$$

$$= 204{,}000\ \text{mm}^3$$

$$\text{or } 7.21 \times 10^{-3}\ \text{ft}^3$$

$$V\ \text{sphere} = \frac{4}{3} \Pi\, r_p^3 = 7.21 \times 10^{-3}$$

$$r_p = 0.120\ \text{ft}$$

$$D_p = 0.240\ \text{ft}$$

The shape factor can now be calculated

$$\phi_s = \text{shape factor} = \frac{\text{surface area of sphere of equivalent volume}}{\text{actual surface area per passage}}$$

$$\phi_s = \frac{\Pi(0.240)\text{ft})^2}{(4)(14\text{ mm})(1000\text{ mm})}\left(\frac{10^6\text{ mm}^2}{\text{m}^2}\right)\left(\frac{.0929\text{ m}^2}{\text{ft}^2}\right)$$

$$= 0.301$$

$$\text{Void fraction} = \varepsilon = 1 - \frac{194\text{ m}^2/\text{m}^3\text{ packed reactor volume}}{601\text{ m}^2/\text{m}^3\text{ catalyst volume}}$$

$$\varepsilon = 1 - 0.323\,\frac{\text{catalyst volume}}{\text{reactor volume}}$$

$$\varepsilon = 0.677\,\frac{\text{void volume}}{\text{reactor volume}}$$

Gas Viscosity, $\mu^o$

This viscosity of a gas mixture can be calculated from[4]

$$\mu^o_{\text{mixture}} = \frac{\Sigma y_i \mu_i (M_i)^{\frac{1}{2}}}{\Sigma y_i (M_i)^{\frac{1}{2}}} \tag{A8-2}$$

where $\mu^o_{\text{mixture}}$ = viscosity of gas mixture, micropoise

$y_i$ = mole fraction of component i in mixture, dimensionless

$\mu_i$ = gas viscosity of pure component i, micropoise

$M_i$ = molecular weight of component i, g/g mole

$\mu_i$ can be calculated from[5]

$$\mu_i = \frac{33.3(M_i T_c)^{\frac{1}{2}}}{V_c^{\,2/3}} \times f\,(1.33T_r) \tag{A8-3}$$

where $T_c$ = critical temperature of component i, °K

$V_c$ = critical volume of component i, $cm^3/g\text{-mole}$

$T_r$ = reduced temperature = ratio of gas temperature to critical temperature ($T/T_c$), dimensionless

$f(T_r)$ = gas viscosity temperature function, dimensionless

Values for $\mu_i$ were calculated using data from Smith & VanNess[6]

| | $T_c$, °k | $T_r$, °k | $V_c, \frac{cm^3}{gmole}$ | $f(1.33T_r)$ | M | $\mu_i, \frac{lb}{ft\text{-}sec}$ |
|---|---|---|---|---|---|---|
| $N_2$ | 126.2 | 5.33 | 89.5 | 3.07 | 28.02 | $2.04 \times 10^{-5}$ |
| $CO_2$ | 304.2 | 2.21 | 94.0 | 1.65 | 44.01 | $2.07 \times 10^{-5}$ |
| $H_2O$ | 647.1 | 1.04 | 56.0 | 0.862 | 18.02 | $1.42 \times 10^{-5}$ |
| $O_2$ | 154.6 | 4.35 | 73.4 | 2.68 | 32.00 | $2.41 \times 10^{-5}$ |

The following data were used with equation (A8-2) to calculate the gas viscosity

| | $y_i$ | $\mu_i, \frac{lb}{ft\text{-}sec}$ | $M_i$ |
|---|---|---|---|
| $N_2$ | 0.728 | $2.04 \times 10^{-5}$ | 28.02 |
| $CO_2$ | 0.132 | $2.07 \times 10^{-5}$ | 44.01 |
| $H_2O$ | 0.095 | $1.42 \times 10^{-5}$ | 18.02 |
| $O_2$ | 0.043 | $2.41 \times 10^{-5}$ | 32.00 |

with the following result

$$\mu^o_{mixture} = 2.00 \times 10^{-5} \frac{lb}{ft\text{-}sec}$$

## Gas Superficial Mass Velocity, G

From the results of the combustion calculation (total mass flow of flue gas) and the reactor sizing calculation (reactor width), the superficial mass velocity can be found.

$$G = \frac{\left(222{,}600\ \frac{\text{lb}}{\text{hr}}\right)\left(\frac{1\ \text{hr}}{3600\ \text{sec}}\right)}{(4.16\text{m})^2\left(10.76\ \frac{\text{ft}^2}{\text{m}^2}\right)} = 0.332\ \frac{\text{lb}}{\text{ft}^2\text{sec}}$$

Modified Reynolds Number, $N'_{Re}$

Using the results of the catalyst characterization, gas mixture viscosity, and mass velocity calculations the modified Reynolds number can be found.

$$N'_{Re} = \frac{D_p G}{\mu^o} = \frac{(0.240\ \text{ft})\left(0.332\ \frac{\text{lb}}{\text{ft}^2\text{-sec}}\right)}{2.00\text{x}10^{-5}\ \frac{\text{lb}}{\text{ft-sec}}} = 3980$$

Knowing the modified Reynolds number the friction factor and exponent can then be determined.[3]

$f_m = 0.7$

$n = 1.97$

Flue Gas Density, $\rho$

Reaction T = 750°F = 1210°R

Volumetric flue gas flow = 73,200 acfm @ 350°F (PedCo)

From combustion calculation: mass flow = $222{,}600\ \frac{\text{lb}}{\text{hr}}$

$$\text{Density} = \frac{\text{mass}}{\text{volume}}$$

$$\rho = \frac{222{,}600\ \frac{\text{lb}}{\text{hr}}}{\left(73{,}200\ \frac{\text{ft}^3}{\text{min}}\right)\left(\frac{1210°\text{R}}{350+460°\text{R}}\right)\left(\frac{60\ \text{min}}{\text{hr}}\right)} = 0.0339\ \frac{\text{lb}}{\text{ft}^3}$$

All of these terms are then substituted into equation (A8-1) to determine the pressure drop.

$$\Delta P = \frac{2f_m G^2 L\ (1-\varepsilon)^{3-n}}{D_p g_c \rho \phi_s^{3-n} \varepsilon^3} \qquad \text{(A8-1)}$$

$$= \frac{2\ (0.7)\left(0.332\ \frac{lb_m}{ft^2 sec}\right)\left[(4.5m)\left(\frac{ft}{.3048m}\right)\right](1-0.677)^{3-1.97}}{(0.240\ ft)\left(32.2\ \frac{lb_m ft}{sec^2 lb_f}\right)\left(0.0339\ \frac{lb_m}{ft^3}\right)(0.301)^{3-1.97}\ (0.677)^3}$$

$$= 30.2\ \frac{lb_f}{ft^2}\left(\frac{ft^2}{144\ in^2}\right)$$

$$\Delta P = 0.210\ psi = 148\ mm\ H_2O$$

Now the energy consumption of the various process steps can be calculated. The energy consuming items considered in this case are

flue gas fan,
liquid $NH_3$ pump,
$NH_3$ vaporization, and
$NH_3$ dilution.

Flue Gas Fan

The gas side horsepower (hp) can be calculated from[7]

$$(hp)_{gas} = 1.57x10^{-4}\ Q\ \Delta P \qquad \text{(A8-4)}$$

$$Q = ft^3/min$$

$$\Delta P = in\ H_2O$$

$$(hp)_{air} = (1.57x10^{-4})\left(73,200\ \frac{ft^3}{min}\right)(148mm\ H_2O)\left(\frac{1\ in}{25.4mm}\right)$$

$$= 67.0\ hp$$

Fan efficiencies typically range from 40 to 70%.[7] If an efficiency of 55% is assumed the shaft hp can be calculated

$$(\text{hp})_{\text{shaft}} = \frac{67.0}{0.55} = 122 \text{ hp}$$

In terms of electrical usage, this is

$$(122 \text{ hp})\left(\frac{0.7457 \text{ kW}}{\text{hp}}\right) = 90.8 \text{ kW}$$

Liquid $NH_3$ Pump

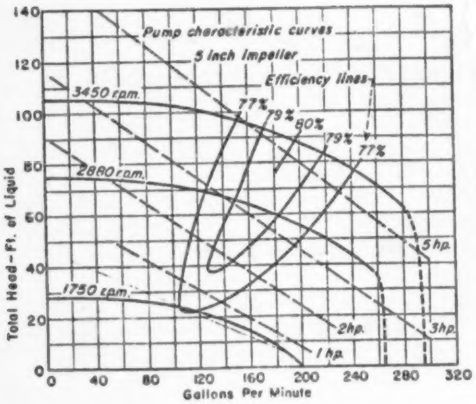
For 90% removal an $NH_3$:NO mole ratio of 0.95 is typical[2]

$$NH_3 \text{ requirement} = \left(188 \frac{\text{lb } NO_x}{\text{hr}}\right)\left(\frac{1 \text{ mole } NO_x}{46 \text{ lb } NO_x}\right)\left(\frac{.95 \text{ mole } NH_3}{\text{mole } NO_x}\right)\left(\frac{17 \text{ lb } NH_3}{\text{lb mole}}\right)$$

$$= 66 \frac{\text{lb } NH_3}{\text{hr}}$$

or 0.21 gpm

The following pump curve[8] indicates that a 0.5 hp centrifugal pump operated at 1750 rpm can supply 28 ft of $NH_3$ head (7.6 psi). This is adequate to transfer sufficient $NH_3$ to the vaporizer.



In terms of electrical usage this amounts to

$$(0.5\ \text{hp})\left(\frac{0.7457\ \text{kW}}{\text{hp}}\right) = 0.373\ \text{kW}$$

$NH_3$ Vaporization

Looking at the worst case for $NH_3$ vaporization, a cold winter day at a Midwest location, the ambient air temperature might be, say, -10°F.

The pressure in the vaporizer is

$$\begin{array}{lrl} & \text{Saturated vapor pressure } (-10^\circ\text{F}) = & 23.7\ \text{psia} \\ + & \text{Pump head} = & \underline{\ 7.6\ \ \ \ \ \ \ } \\ & & 31.3\ \text{psia} \end{array}$$

The normal boiling point of $NH_3$ is -28.0°F.[9] In order to determine the actual boiling point, and thus the heat load on the vaporizer, it is necessary to evaluate it at the higher pressure. This can be accomplished by use of the Clausius-Clapeyron equation.[10]

$$\Delta H_{vap} = -R\frac{d(\ell n\ P^{sat})}{d(^1/T)} \qquad \text{(A8-5)}$$

At -28.0°F and 1 atm, $\Delta H_{vap} = 589.3\ \frac{Btu}{lb}$

The Clausius-Clapeyron equation is used to determine the boiling temperature at the higher pressure of 31.3 psia. Use of the equation assumes a constant $\Delta H_{vap}$, however, this is not strictly true. For this reason, two iterations are calculated. The first uses $\Delta H_{vap}$ at 14.7 psia and calculates an elevated boiling temperature at P = 31.3 psia. The $\Delta H_{vap}$ at this temperature is found from thermodynamic tables. Then the two $\Delta H_{vap}$ values are averaged to determine a pseudo constant $\Delta H_{vap}$. This value is then used in the second iteration to determine a new and more accurate elevated boiling temperature.

Solving equation (A8-5) gives

$$\ell n\ \frac{P_2}{P_1} = -\ \frac{\Delta H_{vap}}{R}\left(\frac{1}{T_2} - \frac{1}{T_1}\right) \qquad \text{(A8-6)}$$

The data used to calculate $T_2$ are

$P_1 = 14.7$ psia

$P_2 = 31.3$ psia

$T_1 = -28.0°F + 460 = 432°R$

$\Delta H_{vap} = 589.3\ \frac{Btu}{lb}$

$R = 1.986\ \frac{Btu}{lb\ mole\ °R}$

Solution of equation (A8-6) for $T_2$ gives

$T_2 = 462°R = 2°F$

$\Delta H_{vap}$ at 2°F is $567.3\ \frac{Btu}{lb}$.[9]

The two $\Delta H_{vap}$ values are averaged to obtain a pseudo constant $\Delta H_{vap}$

$$\overline{\Delta H}_{vap} = \frac{589.3 + 567.3}{2} = 578.3\ \frac{\text{Btu}}{\text{lb}}$$

Solving equation (A8-6) a second time using this new $\Delta H_{vap}$ indicates a boiling temperature of 3°F. At this temperature, $\Delta H_{vap}$ is 566.5 $\frac{\text{Btu}}{\text{lb}}$.

Now the energy requirement for $NH_3$ vaporization can be calculated. The heat capacity of $NH_3$ at this temperature is[11]

$$C_p\ [NH_3(\ell)] = 1.10\ \frac{\text{Btu}}{\text{lb°F}}$$

$$Q_{vaporization} = \left(66\ \frac{\text{lb } NH_3}{\text{hr}}\right)\left[\left(1.10\ \frac{\text{Btu}}{\text{lb°F}}\right)[3°\text{F}-(-10°\text{F})] + 566.5\ \frac{\text{Btu}}{\text{lb}}\right]$$

$$= 38{,}300\ \frac{\text{Btu}}{\text{hr}}$$

$NH_3$ Dilution

The $NH_3$ is diluted with 30 psig steam prior to injection. A 5:1 mole ratio is used.[12] The heat of vaporization of 30 psig steam is 929.0 $\frac{\text{Btu}}{\text{lb}}$.[13]

$$\text{Mass of steam injected} = \left(66\ \frac{\text{lb } NH_3}{\text{hr}}\right)\left(\frac{\text{mole } NH_3}{17.0\text{ lb } NH_3}\right)\left(\frac{5\text{ moles steam}}{\text{mole } NH_3}\right)$$

$$\left(\frac{18\text{ lb steam}}{\text{mole steam}}\right) = 350\ \frac{\text{lb steam}}{\text{hr}}$$

$$Q = (350\text{ lb})\left(\frac{929.0\text{ Btu}}{\text{lb}}\right) = 325{,}000\ \frac{\text{Btu}}{\text{hr}}$$

Summarizing the energy consuming steps,

| Item | Energy usage | Btu/hr | $MW_{thermal}$ |
|---|---|---|---|
| Flue gas fan | 90.8 $kW_{elec}$ | 908,000 | 0.266 |
| Liquid $NH_3$ pump | 0.373 $kW_{elec}$ | 3,730 | 0.00109 |
| $NH_3$ vaporization | 38,300 Btu/hr | 38,300 | 0.0112 |
| $NH_3$ dilution | 325,000 Btu/hr | 325,000 | 0.0953 |
| Total | | 1,275,030 | 0.374 |

The example calculation is continued below to show the methods used to arrive at annual cost figures for $NO_x$ control systems. The case is the Parallel Flow SCR system applied to the Pulverized Coal standard boiler and operated at the stringent level of control. The material balance and process flow diagram, as it appears in the Appendix, are presented in Figure A8-1 on the following page. First, each of the pieces of equipment, in succession, will be sized (including any necessary design calculations) and the F.O.B. costs determined. These results will then be utilized to determine the installed costs. Then, the direct operating costs are calculated. These costs are combined via the recommended format using the appropriate load factor to arrive at the annual costs.

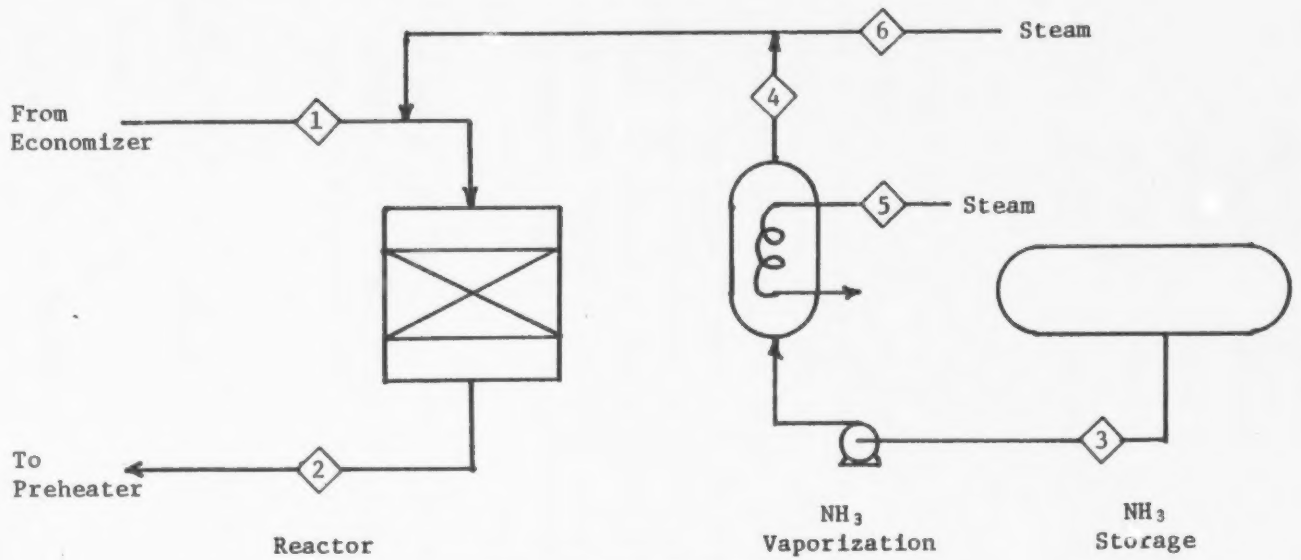
From Figure A8-1, the $NH_3$ flow is shown to be 1.76 $\frac{\text{kg-mole}}{\text{hr}}$. Assuming the plant maintains a 15-day storage supply (large enough to survive delay in deliveries due to bad weather, strikes, etc.), the required $NH_3$ storage tankage is determined.

$$\text{Gal Storage} = \left(1.76\ \frac{\text{kg-mole}}{\text{hr}}\right)\left(\frac{17.0\ \text{kg}}{\text{kg-mole}}\right)\left(\frac{\text{lb}}{.454\ \text{kg}}\right)\left(\frac{\text{ft}^3}{39.0\ \text{lb}}\right)^{17}\left(\frac{7.48\ \text{gal}}{\text{ft}^3}\right)\left(\frac{24\ \text{hr}}{\text{day}}\right)$$

$$\times\ (15\ \text{day supply}) \qquad \text{(A8-7)}$$

$$= 4600\ \text{gal}$$

$$\text{F.O.B. Equipment Cost} = \text{Base Cost (Unit Cost)}^{\text{Exponential Factor}} \times \text{Pressure Factor}$$


Steam
From
Economizer
1
6
4
5
Steam
3
To
Preheater
2
Reactor
NH₃
Vaporization
NH₃
Storage

| | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| T, °K | 648 | 648 | 283 | 289 | 429 | 408 |
| P, Pa | 98,600 | 97,150 | 615,000 | 752,000 | 552,000 | 310,000 |
| $N_2$ | 2495 | 2497 | - | - | - | - |
| $CO_2$ | 453 | 453 | - | - | - | - |
| $H_2O$ | 325 | 336 | - | - | 0.98 | 8.8 |
| $O_2$ | 149 | 149 | - | - | - | - |
| $NO_x$ | 1.86 | 0.19 | - | - | - | - |
| $SO_x$ | 1.69 | 1.69 | - | - | - | - |
| $NH_3$ | - | 0.03 | 1.76 | 1.76 | - | - |

Figure A8-1. Material balance.
Pulverized Coal
Parallel Flow SCR
Stringent Control

$$\text{F.O.B. NH}_3\text{ Tank Cost (mid-1970)}^{15} = 10{,}000\left(\frac{\text{gal}}{10{,}000}\right)^{0.7} \times 1.38 \qquad \text{(A8-8)}$$

$$= 10{,}000\left(\frac{4{,}600}{10{,}000}\right)^{0.7} \times 1.38$$

$$= \$8{,}000$$

Next, the two liquid $NH_3$ pumps (one for a spare in case of failure) are examined. The volumetric liquid $NH_3$ flow rate can be determined from:

$$\text{Liq. NH}_3\text{ Vol. Flow} = \left(1.76\ \frac{\text{kg-mole}}{\text{hr}}\right)\left(\frac{17.0\text{ kg}}{\text{kg-mole}}\right)\left(\frac{\text{lb}}{.454\text{ kg}}\right) \times \left(\frac{\text{ft}^3}{39.0\text{ lb}}\right)\left(\frac{7.48\text{ gal}}{\text{ft}^3}\right)\left(\frac{\text{hr}}{60\text{ min}}\right) \qquad \text{(A8-9)}$$

$$= 0.21\text{ GPM}$$

At this flow rate, a 0.5 hp centrifugal pump is adequate.[8]

F.O.B. Equipment Cost = H Pumps x Base Cost

$$\text{F.O.B. Pump Cost (mid-1970)}^{16} = 2\text{ pumps} \times \frac{\$300}{\text{pump}}$$

$$= \$600$$

To size the $NH_3$ vaporizer the sensible heat and the heat of vaporization for $NH_3$ is required. For the worst case at -10°F and 31 psia tank pressure, the heat required = $581\ \frac{\text{Btu}}{\text{lb NH}_3}$.

The heat load on the vaporizer is shown to be:

$$Q = \left(1.76\ \frac{\text{kg-mole}}{\text{hr}}\right)\left(\frac{17.0\text{ kg}}{\text{kg-mole}}\right)\left(\frac{\text{lb}}{.454\text{ kg}}\right)\left(\frac{581\text{ BTU required}}{\text{lb NH}_3\text{ vaporized}}\right)$$

$$= 38{,}300\ \frac{\text{BTU}}{\text{hr}}$$

The heat transfer area required for the $NH_3$ vaporizer can be calculated from:

$$A = \frac{qF}{U_c(T_n - T)} \quad ^{17} \qquad \text{(A8-10)}$$

where A = heat transfer surface area, $ft^2$

q = heat transferred, BTU/hr

F = safety factor (assume 2.0)

$U_c$ = heat transfer coefficient, BTU/hr $ft^2$ °F

$T_n$ = temperature of heat medium, °F
(800 psia steam = 312°F)

T = fluid temperature, °F
(worst case: $NH_3$ = -10°F)

$$A = \frac{(38{,}300 \frac{BTU}{hr})(2.0)}{\left(300 \frac{BTU}{hr\text{-}ft^2\,°F}\right)^{18}\left(312 - (-10)°F\right)}$$

$$= 0.79\ ft^2$$

The smallest commercially available doube pipe heat exchanger is 1 $ft^2$ and the F.O.B. Vaporizer Cost (mid-1970) = \$300.[19]

Next, the reactor size is determined from the volumetric flue gas flow rate and the reactor space velocity.

$$\text{Volume Flue Gas} = \left(73{,}200 \frac{ft^2}{min}\right)\left(\frac{492°R}{810°R}\right)\left(\frac{0.0283\ Nm^3}{scf}\right)\left(\frac{60\ min}{hr}\right)$$

$$= 75{,}500 \frac{Nm^3}{hr} \qquad \text{(A8-11)}$$

$$\text{Catalyst Volume}^{20} = \frac{75{,}500\ Nm^3/hr}{3000\ hr^{-1}\ \text{space velocity}} = 25.2\ m^3 = 889\ ft^3$$

(A8-12)

Reactor Volume[21] = $(25.2 \text{ m}^3 \text{ catalyst}) \times$

$$\left(\frac{601 \text{ m}^2 \text{ surface area/m}^3 \text{ catalyst}}{194 \text{ m}^2 \text{ surface area/m}^3 \text{ packed volume}}\right)$$

$$= 78.0 \text{ m}^3 \text{ packed volume} \qquad \text{(A8-13)}$$

Reactor Length[21] = $4.5 \text{ m} \left(\frac{\text{ft}}{.3048\text{m}}\right) = 14.8 \text{ ft}$

Square Reactor Volume = $78.0 \text{ m}^3 = W^2L = 4.5\ W^2$

$$W^2 = \frac{78.0 \text{ m}^3}{4.5 \text{ m}^3} = 17.3 \text{ m}^2$$

$$W = 4.16 \text{ m} = 13.6 \text{ ft}$$

F.O.B. Reactor Material Cost (mid-1970) = \$16,000[22]

Finally, the draft fan motor drive must be determined.

$$\text{Motor hp.} = \frac{0.000157\ Q\ \Delta p}{55\%\ \text{efficiency}}^{7} \qquad \text{(A8-4)}$$

$$= \frac{0.000157 \left(73{,}200 \frac{\text{ft}}{\text{min}}\right) (148 \text{ mmH}_2\text{O}) \left(\frac{\text{in}}{25.4 \text{ mm}}\right)}{.55}$$

$$= 122 \text{ hp}$$

Motor Drive - F.O.B. Motor Cost (mid-1970)[23] $= 5800\left(\frac{\text{hp}}{70}\right)^{0.77}$ (A8-14)

$$= 5800\left(\frac{122}{70}\right)^{0.77}$$

$$= \$8{,}900$$

Each piece of equipment is factored by its respective escalation index to give a 1978 F.O.B. cost and an 8 percent freight charge for delivery is added to it.[24] The direct installation costs are determined by the appropriate factor multiplied times the 1970 F.O.B. equpment cost and that category's respective escalation index. These analyses for each equipment item are presented on the following pages.

## $NH_3$ Storage Tank

F.O.B. Equipment Cost (mid-1970) = $10{,}000 \left(\frac{\text{gal.}}{10{,}000}\right)^{0.7} \times 1.38$ = \$ 8,000

### June - 1978 Costs

| | | |
|---|---|---|
| Equipment Cost (1978) | | |
| Basic equipment | = F.O.B. 1970 x Escalation Index<br>= \$8000 x 1.91 | = 15,300 |
| Freight | = 0.08 x Basic Equipment | = 1,200 |
| Required auxiliaries | | = N/A |
| | Total Equipment Cost | = 16,500 |
| Installation Costs, Direct[15] | = F.O.B. 1970 x Installation Cost Fraction x Escalation Index | |
| Foundation and supports | = F.O.B. 1970 x 0.080 x 2.11 | = 1,400 |
| Ductwork | | = N/A |
| Stack | | = N/A |
| Piping | = F.O.B. 1970 x 0.153 x 2.02 | = 2,500 |
| Insulation | | = N/A |
| Painting | = F.O.B. 1970 x 0.012 x 2.11 | = 200 |
| Electrical | = F.O.B. 1970 x 0.007 x 1.70 | = 100 |
| Instruments | = F.O.B. 1970 x 0.118 x 1.63 | = 1,500 |
| Installation Labor | = F.O.B. 1970 x 0.352 x 1.37 | = 3,900 |
| | Total Installation Cost | = 9,600 |
| | Total Direct Cost | = \$26,100 |

## Liquid $NH_3$/Pumps

2 - 0.5 hp Centrifugal Pumps (1 spare)

Pump and motor - F.O.B. Equipment Cost (mid-1970) = \$300 x 2 = \$ 600

### June - 1978 Costs

| | | |
|---|---|---|
| Equipment Cost (1978) | | |
| Basic equipment | = F.O.B. 1970 x Escalation Index | |
| | = \$600 x 2.08 | = 1,250 |
| Freight | = 0.08 x Basic equipment | = 100 |
| Required auxiliaries | | = N/A |
| | Total Equipment Cost | = 1,350 |
| Installation Costs, Direct[16] | = F.O.B. 1970 x Installation Cost Fraction x Escalation Index | |
| Foundation and supports | = F.O.B. 1970 x 0.039 x 2.11 | = 50 |
| Ductwork | | = N/A |
| Stack | | = N/A |
| Piping | = F.O.B. 1970 x 0.293 x 2.02 | = 360 |
| Insulation | = F.O.B. 1970 x 0.028 x 2.11 | = 40 |
| Painting | = F.O.B. 1970 x 0.008 x 2.11 | = 10 |
| Electrical | = F.O.B. 1970 x 0.303 x 1.70 | = 310 |
| Instruments | = F.O.B. 1970 x 0.029 x 1.63 | = 30 |
| Installation labor | = F.O.B. 1970 x 0.679 x 1.37 | = 560 |
| | Total Installation Cost | 1,360 |
| | Total Direct Cost | \$2,710 |

## $NH_3$ Vaporizer

1 $ft^2$ Double-Pipe Heat Exchanger
(minimum size available)

Vaporizer - F.O.B. Equipment Cost (mid-1970) = \$300

| | June - 1978 Costs | |
|---|---|---|
| Equipment Costs (1978) | | |
| Basic equipment | = F.O.B. 1970 x Escalation Index<br>= \$300 x 1.91 | = \$ 570 |
| Freight | = 0.08 x Basic equipment | = 50 |
| Required auxiliaries | | = N/A |
| | Total Equipment Cost | = 620 |
| Installation Costs, Direct[19] | = F.O.B. 1970 x Installation Cost Fraction x Escalation Index | |
| Foundation and supports | = F.O.B. 1970 x 0.038 x 2.11 | = 50 |
| Ductwork | | = N/A |
| Stack | | = N/A |
| Piping | = F.O.B. 1970 x 0.213 x 2.02 | = 270 |
| Insulation | = F.O.B. 1970 x 0.022 x 2.11 | = 30 |
| Painting | = F.O.B. 1970 x 0.002 x 2.11 | = --- |
| Electrical | = F.O.B. 1970 x 0.010 x 170 | = 10 |
| Instruments | = F.O.B. 1970 x 0.048 x 1.63 | = 50 |
| Installation labor | = F.O.B. 1970 x 0.467 x 1.37 | = 400 |
| | Total Installation Cost | = 810 |
| | Total Direct Cost | = \$1,430 |

## Reactor

| | | |
|---|---|---|
| Reactor - F.O.B. Equipment Cost (mid-1970): | Material | = $ 16,000 |
| : | Fabricated (2x Material) | = 32,000 |

### June - 1978 Costs

| | | |
|---|---|---|
| Equipment Cost (1978) | | |
| Basic equipment | = F.O.B. 1970 x Escalation Index<br>= $32,000 x 1.91 | = 61,100 |
| Required auxiliary:Catalyst | = $212 x $ft^3$ catalyst | = 188,000 |
| Freight | = 0.08 x Basic equipment | = 4,890 |
| | Total Equipment Cost | = 254,000 |
| Installation Costs, Direct[22] | = F.O.B. 1970 x Installation Cost Fraction x Escalation Index | |
| Foundation and supports | = F.O.B. 1970 x 0.176 x 2.11 | = 11,800 |
| Ductwork | | = N/A |
| Stack | | = N/A |
| Piping | = F.O.B. 1970 x 0.595 x 2.02 | = 38,400 |
| Insulation | = F.O.B. 1970 x 0.080 x 2.11 | = 5,400 |
| Painting | = F.O.B. 1970 x 0.013 x 2.11 | = 880 |
| Electrical | = F.O.B. 1970 x 0.049 x 1.70 | = 2,600 |
| Instruments | = F.O.B. 1970 x 0.114 x 1.63 | = 6,000 |
| Installation labor | = [F.O.B. 1970 x 0.972 x 1.37 + (Catalyst x 0.10)] | = 61,400 |
| | Total Installation Cost | = 126,480 |
| | Total Direct Cost | $380,470 |

## Draft Fan Motor Drive

Motor - F.O.B. Equipment Cost (mid-1970) = $5{,}800\left(\frac{hp}{70}\right)^{0.77}$ = $ 8,900

### June - 1978 Costs

| | | | |
|---|---|---|---|
| Equipment Cost (1978) | | | |
| Basic equipment | = F.O.B. 1970 x Escalation Index<br>= $8900 x 2.08 | = | 18,500 |
| Freight | = 0.08 x Basic equipment | = | 1,500 |
| Required auxiliaries | | = | N/A |
| | Total Equipment Cost | | 20,000 |
| Installation Costs, Direct[25] | = F.O.B. 1970 x Installation Cost Fraction x Escalation Index | | |
| Foundation and supports | = F.O.B. 1970 x 0.043 x 2.11 | = | 820 |
| Ductwork | | = | N/A |
| Stack | | = | N/A |
| Piping | = F.O.B. 1970 x 0.141 x 2.02 | = | 2,500 |
| Insulation | = F.O.B. 1970 x 0.005 x 2.11 | = | 90 |
| Painting | | = | N/A |
| Electrical | = F.O.B. 1970 x 0.068 x 1.70 | = | 1,000 |
| Instruments | = F.O.B. 1970 x 0.013 x 1.63 | = | 190 |
| Installation labor | = F.O.B. 1970 x 0.295 x 1.37 | = | 3,600 |
| | Total Installation Cost | = | 8,200 |
| | Total Direct Cost | = | $28,200 |

The direct operating costs are shown below.

Ammonia

$$\left(1.76\ \frac{\text{kg mole}}{\text{hr}}\right)\left(\frac{\$130}{\text{ton}}\right)\left(\frac{1\ \text{ton}}{2000\ \text{lb}}\right)\left(\frac{17\ \text{kg}}{\text{kg mole}}\right)\left(\frac{\text{lb}}{454\ \text{g}}\right)\left(\frac{1000\ \text{g}}{\text{kg}}\right)(8760\ \text{hr})(0.6)$$

$= \$22,535$

Electricity

$$(0.268\ \text{MW}_t)\left(\frac{25.8\ \text{mills}}{\text{kwh}}\right)\left(\frac{\$1}{1000\ \text{mills}}\right)\left(\frac{342\ \text{kw}}{\text{MW}_t}\right)(8760\ \text{hr})(0.6)$$

$= \$12,429$

Steam

$$\left(\frac{9.78\ \text{kg mole}}{\text{hr}}\right)\left(\frac{18\ \text{kg}}{\text{kg mole}}\right)\left(\frac{1000\ \text{g}}{\text{kg}}\right)\left(\frac{1\ \text{lb}}{454\ \text{g}}\right)\left(\frac{\$3.50}{1000\ \text{lb}_m}\right)(8760\ \text{hr})(0.6)$$

$= \$7,133$

The individual equipment costs and installation costs are summed and the totals entered in Table A8-1. The direct operating costs are entered in Table A8-2.

## TABLE A8-1. CAPITAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Equipment cost | | | |
| Basic equipment (includes freight) | 104,470 | | |
| Required auxiliaries | 188,000 | | |
| Total equipment cost | 292,470 | | |
| Installation costs, direct | | | |
| Foundations and supports | 14,120 | | |
| Piping | 44,030 | | |
| Insulation | 5,560 | | |
| Painting | 1,090 | | |
| Electrical | 4,020 | | |
| Instruments | 7,770 | | |
| Installation labor | 69,860 | | |
| Total installation cost | 146,450 | | |
| Total Direct Costs (equipment + installation) | | 438,920 | |
| Installation costs, indirect | | | |
| Engineering | 43,892 | | |
| Construction and field expense | 43,892 | | |
| Construction fees | 43,892 | | |
| Start-up | 8,778 | | |
| Performance tests | 2,000 | | |
| Total Indirect Costs | | 142,454 | |
| Contingencies | | 116,275 | |
| Total Turnkey Costs (direct+indirect+contingencies) | | 697,649 | |
| Land | | 1,744 | |
| Working capital | | 54,288 | |
| GRAND TOTAL (turnkey + land + working capital) | | | $753,620 |

TABLE A8-2. ANNUAL COSTS

Boiler type: Pulverized Coal
Fuel: Low Sulfur Western Coal
Control technique: Parallel Flow SCR
Control level: Stringent

| | | | |
|---|---|---|---|
| Direct costs | | | |
| Direct labor | 17,279 | | |
| Maintainence labor | 31,569 | | |
| Materials | 12,884 | | |
| Catalyst | 113,081 | | |
| Electricity | 12,417 | | |
| Steam | 7,133 | | |
| Ammonia | 22,535 | | |
| Total direct cost | | 216,900 | |
| Overhead | | | |
| Payroll | 5,184 | | |
| Plant | 16,050 | | |
| Total overhead cost | | 21,234 | |
| Capital Charges | | | |
| G&A, taxes & ins. | 27,906 | | |
| Capital recovery | 91,720 | | |
| Total capital charges | | 119,626 | |
| TOTAL ANNUALIZED COSTS | | | $357,760 |

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19. Guthrie, K.M., *op.cit.*, pp. 144, 145.

20. Ando, J., *op.cit.*, p. 3-31.

21. *Ibid.*, p. 3-30.

22. Guthrie, K.M., *op.cit.*, pp. 150-154.

23. Woods, Donald R. Financial Decision Making in the Process Industry. Prentice-Hall. 1975. p. 301.

24. Guthrie, K.M. "Capital Cost Estimating." Chemical Engineering. March 24, 1969. p. 122.

25. *Ibid.*, p. 174.